LEGIBILITY NOTICE

A major purpose of the Technical Information Center is to provide the broadest dissemination possible of information contained in DOE's Research and Development Reports to business, industry, the academic community, and federal, state and local governments.

Although a small portion of this report is not reproducible, it is being made available to expedite the availability of information on the research discussed herein.

1

EPA--600/7-83-029
DE84 900344

EPA-600/7-83-029
TVA/OP/EDT-82/6
May 1983

Economic Evaluation of Limestone and Lime Flue Gas Desulfurization Processes

by

T.A. Burnett, C.D. Stephenson,
F.A. Sudhoff, and J.D. Veitch

TVA, Office of Power
Division of Energy Demonstrations and Technology
Muscle Shoals, Alabama 35660

EPA Interagency Agreement No. 79-D-X0511

EPA Project Officer: J. David Mobley

Industrial Environmental Research Laboratory
Office of Environmental Engineering and Technology
Research Triangle Park, NC 27711

NOTICE
PORTIONS OF THIS REPORT ARE ILLEGIBLE.
It has been reproduced from the best available copy to permit the broadest possible availability.

[illegible]UTION OF THIS DOCUMENT IS [illegible]

Prepared for

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Research and Development
Washington, DC 20460

MASTER

DISCLAIMER

This report was prepared by the Tennessee Valley Authority and has been reviewed by the Office of Energy, Minerals, and Industry, U.S. Environmental Protection Agency, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the Tennessee Valley Authority or the U.S. Environmental Protection Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

ABSTRACT

The preliminary-grade economics (accuracy: -15%, +30%) of various alternative limestone scrubbing options (absorber type, with and without forced oxidation, and with and without adipic acid enhancement) are examined using the current design and economic premises established for the continuing series of economic evaluations performed by TVA for EPA. The economics are projected using the Shawnee lime/limestone computer model, which is based on long-term operating data from the EPA Alkali Scrubbing Test Facility at the TVA Shawnee Steam Plant near Paducah, Kentucky. The capital investment for the base-case limestone scrubbing process (500 MW, 3.5% sulfur coal, 1979 NSPS, spray tower, forced oxidation, landfill) is $206/kW. The first-year and levelized annual revenue requirements are 10.59 and 15.09 mills/kWh, respectively. Costs for the equivalent limestone scrubbing process using a Turbulent Contact Absorber (TCA) are lower while those for the venturi - spray tower absorber are higher. The forced-oxidation landfill disposal option has a lower capital investment than the unoxidized pond disposal option for all cases studied; however, the first-year and levelized annual revenue requirements are slightly higher for the forced-oxidation landfill process for most coal applications. For the spray tower limestone process to achieve a specified SO_2 removal efficiency, it is more economical to increase the limestone stoichiometry and minimize the absorber L/G. The use of adipic acid or possibly dibasic acid (DBA) as an additive to enhance SO_2 removal in the limestone scrubbing process is an economically attractive option. The use of adipic acid remains economically attractive even if both a high unit cost and a high degradation factor for adipic acid are assumed.

CONTENTS

Abstract ii
Figures vii
Tables ix
Abbreviations and Conversion Factors xii

Executive Summary S-1

Introduction 1

Background 3
  Early Limestone and Lime FGD Developments 3
  Chemistry of Absorption 5
    Simplified SO_2-CO_2 Systems 5
    Dissolution and Precipitation Reactions 8
    Overall Chemistry 11
    Other Ionic Species 11
    Lime Processes 11
  Oxidation of Sulfite 12
    Catalysts and Inhibitors 13
    Forced Oxidation FGD 15
  Additive-Enhanced Scrubbing Liquids 16
    Magnesium-Enhanced Scrubbing Liquids 16
    Buffer Additives 18
  Scale Formation 20
  Waste Treatment and Disposal 21
  Absorbers 24
    Venturi 24
    Spray Towers 25
    Mobile-Bed Absorbers 26
    Comparison 27
  Construction Materials 28
  The Shawnee Lime/Limestone Scrubbing Model 30

Design and Economic Premises 33
  Design Premises 33
    Coal 33
    Power Plant 33
    Flue Gas Compositions 35
    Environmental Regulations 35
    Particulate Matter 42
    Flue Gas Desulfurization 42
    Solids Disposal 44

Economic Premises 45
  Schedule and Cost Factors 45
  Capital Cost Estimates 45
  Annual Revenue Requirements 51

Systems Estimated 55
 Major Processing Areas 55
  Materials Handling Area 55
  Feed Preparation Area 58
  Gas Handling Area 58
  SO_2 Absorption Area 59
  Oxidation Area 61
  Reheat Area 61
  Solids Separation and Disposal Areas 62
  Configurations 62
 Basic Systems 64
  Limestone Process with a TCA, Without Forced Oxidation, and Pond Disposal (Case 6) 64
  Limestone Process with a Spray Tower Absorber, Forced Oxidation, and Landfill Disposal (Case 15) 73
  Limestone Process with a Venturi - Spray Tower Absorber, Forced Oxidation, and Landfill Disposal (Case 22) 84
  Lime Process with a Venturi - Spray Tower Absorber, MgO Addition, Bleedstream Forced Oxidation, and Landfill Disposal (Case 52) . . 84

Results 105
 Cost Summaries and Detailed Comparisons 105
  Comparison of Limestone Processes With and Without Forced Oxidation 105
  Comparison of Limestone Processes With and Without Additives . . . 113
  Power Unit Size and Sulfur Contents 118
  Comparisons by Type of Absorber 118
  Spray Tower Absorber Design Conditions 124
  Partial Scrubbing Versus Full Scrubbing 124
  Full-Scrubbing Processes at 85%, 90%, and 95% SO_2 Removal 127
  No Reheat 127
  Comparisons of Limestone and Lime Processes 127
  Cost Sensitivities 133
 Comparison of Previous and Current TVA Limestone FGD Economics . . . 144
  Design Premises 146
  Economic Premises 147
  Cost Comparison 147

Conclusions 153

References 155

Appendix A A-1
Appendix B B-1
Appendix C C-1
Appendix D D-1

FIGURES

Number		Page
S-1	Cost comparison of 500-MW limestone processes with and without forced oxidation at different coal sulfur contents . . .	S-11
S-2	Cost comparison of limestone processes with and without forced oxidation at different absorber stoichiometries . . .	S-16
S-3	Cost comparison of 500-MW limestone forced-oxidation processes with partial or full scrubbing at different coal sulfur contents	S-17
S-4	Cost comparison of 500-MW limestone forced-oxidation processes with and without adipic acid at different SO_2 removal efficiencies	S-19
S-5	Sensitivity of the limestone forced-oxidation process costs to changes in inflation and discount rate	S-23
S-6	Sensitivity of limestone forced-oxidation process costs to changes in labor and maintenance costs	S-24
S-7	Sensitivity of lime process costs to the cost of lime	S-26
S-8	Sensitivity of limestone adipic acid forced-oxidation process costs to the cost of adipic acid	S-27
S-9	Sensitivity of limestone adipic acid forced-oxidation process costs to the adipic acid degradation rate	S-28
1	Relationship of sulfur dioxide and carbonate species in solution as a function of pH	7
2	Controlled SO_2 emission requirements for 1979 NSPS	41
3	Pond plan and dike construction details	46
4	Landfill plan and construction details	47
5	Solids separation area, flow diagram	63
6	Limestone process with a TCA, without forced oxidation, and pond disposal, flow diagram	66
7	Limestone process with a TCA, without forced oxidation, and pond disposal, plan and elevation	68
8	Limestone process with a spray tower absorber, forced oxidation, and landfill disposal, flow diagram	74
9	Limestone process with a spray tower absorber, forced oxidation, and landfill disposal, plan and elevation	77
10	Limestone process with a venturi - spray tower absorber, forced oxidation, and landfill disposal, flow diagram	85
11	Limestone process with a venturi - spray tower absorber, forced oxidation, and landfill disposal, plan and elevation .	88
12	Lime process with a venturi - spray tower absorber, MgO addition, bleedstream forced oxidation, and landfill disposal, flow diagram	95

FIGURES (continued)

Number Page

13 Lime process with a venturi - spray tower absorber, MgO addition, bleedstream forced oxidation, and landfill disposal, plan and elevation 98
14 Cost comparison of 500-MW limestone processes with and without forced oxidation at different coal sulfur contents . 112
15 Cost comparison of limestone forced-oxidation processes with and without additives at different coal sulfur contents . . . 116
16 Cost comparison of power unit size and coal sulfur content for limestone forced-oxidation processes 121
17 Cost comparison of limestone processes with and without forced oxidation at different absorber stoichiometries . . . 125
18 Cost comparison of 500-MW limestone forced-oxidation processes with partial or full scrubbing at different coal sulfur contents 126
19 Cost comparison of 500-MW limestone forced-oxidation processes with and without adipic acid at different SO_2 removal efficiencies 128
20 Sensitivity of limestone forced-oxidation process costs to changes in inflation and discount rate 136
21 Sensitivity of limestone forced-oxidation process costs to changes in labor and maintenance costs 137
22 Annual revenue requirement variation with maintenance costs for limestone TCA and spray tower landfill processes 139
23 Sensitivity of lime process costs to the cost of lime 140
24 Sensitivity of limestone adipic acid forced-oxidation process costs to the cost of adipic acid 143
25 Sensitivity of limestone adipic acid forced-oxidation process costs to the adipic acid degradation rate 145
26 Stepwise conversion of limestone process costs from old to new premises and technology 151

TABLES

Number | | Page
S-1 Processes Evaluated S-4
S-2 Capital Investment and Annual Revenue Requirements for Limestone Processes Without Forced Oxidation S-6
S-3 Capital Investment and Annual Revenue Requirements for Limestone Processes With Forced Oxidation S-7
S-4 Capital Investment and Annual Revenue Requirements for Limestone Processes With Forced Oxidation and Additives . . . S-8
S-5 Capital Investment and Annual Revenue Requirements for Lime Processes S-9
S-6 Cost Comparison of Processes With and Without Forced Oxidation S-10
S-7 Cost Comparison of Processes With and Without Additives . . . S-12
S-8 Cost Comparison by Absorber Type S-15
S-9 Capital Investment and Annual Revenue Requirements for Limestone Processes With and Without Reheat S-20
S-10 Cost Comparison of Limestone and Lime Processes S-21
S-11 Cost Comparison of Adipic Acid and DBA Enhanced Limestone Processes S-25
1 Coal Compositions 34
2 Power Unit Operating Time and Heat Rate 35
3 Flue Gas Composition for 5% Sulfur Eastern Bituminous Coal 36
4 Flue Gas Composition for 3.5% Sulfur Eastern Bituminous Coal 37
5 Flue Gas Composition for 2% Sulfur Eastern Bituminous Coal 38
6 Flue Gas Composition for 0.7% Sulfur Western Bituminous Coal 39
7 1979 Revised NSPS Emission Standards 40
8 Premise Coal Emission Standards 41
9 Raw Material Characteristics 44
10 Cost Indexes and Projections 49
11 Cost Factors 48
12 Indirect Capital Cost Factors 50
13 Process Combinations 56
14 Processes Evaluated 57
15 Absorber Operating Conditions 59
16 Summary of Operating Conditions 65

TABLES (continued)

Number Page

17 Limestone Process With a TCA, Without Forced Oxidation, and Pond Disposal Material Balance 67

18 Limestone Process With a TCA, Without Forced Oxidation, and Pond Disposal Equipment List 69

19 Limestone Process With a Spray Tower Absorber, Forced Oxidation, and Landfill Disposal Material Balance 75

20 Limestone Process With a Spray Tower Absorber, Forced Oxidation, and Landfill Disposal Equipment List 78

21 Limestone Process With a Venturi - Spray Tower Absorber, Forced Oxidation, and Landfill Disposal Material Balance . . 86

22 Limestone Process With a Venturi - Spray Tower Absorber, Forced Oxidation, and Landfill Disposal Equipment List . . . 89

23 Lime Process With a Venturi - Spray Tower Absorber, MgO Addition, Bleedstream Forced Oxidation, and Landfill Disposal Material Balance 96

24 Lime Process With a Venturi - Spray Tower Absorber, MgO Addition, Bleedstream Forced Oxidation, and Landfill Disposal Equipment List 99

25 Capital Investment and Annual Revenue Requirements for Limestone Processes Without Forced Oxidation 106

26 Capital Investment and Annual Revenue Requirements for Limestone Processes With Forced Oxidation 107

27 Capital Investment and Annual Revenue Requirements for Limestone Processes With Forced Oxidation and Additives . . . 108

28 Capital Investment and Annual Revenue Requirements for Lime Processes 109

29 Capital Investment Comparison of Processes With and Without Forced Oxidation 110

30 Annual Revenue Requirements Comparison of Processes With and Without Forced Oxidation 111

31 Capital Investment Comparison for Processes With and Without Additives 114

32 Annual Revenue Requirements Comparison for Processes With and Without Additives 115

33 Capital Investment Comparison at Various Power Unit Sizes . . 119

34 Annual Revenue Requirements Comparison at Various Power Unit Sizes 120

35 Capital Investment Comparison by Absorber and Process Type . 122

36 Annual Revenue Requirements Comparison by Absorber and Process Type 123

37 Capital Investment and Annual Revenue Requirements for Limestone Processes With and Without Reheat 129

38 Capital Investment Comparison for Limestone and Lime Processes 130

39 Annual Revenue Requirements Comparison for Limestone and Lime Processes 131

TABLES (continued)

Number **Page**

40 Cost Distribution as a Percentage of Total First-Year Annual Revenue Requirements 135
41 Cost Sensitivities 134
42 Cost Comparison of Adipic-Acid- and DBA-Enhanced Limestone Processes 141
43 Sensitivity of Limestone Process Costs to Adipic Acid Cost and Consumption 142
44 Comparison of Old and New Premise Conditions for the Limestone Process 148
45 Stepwise Conversion of Limestone Process Costs from Old to New Premises and Technology 150

ABBREVIATIONS AND CONVERSION FACTORS

ABBREVIATIONS

aft^3/min	actual cubic feet per minute
Btu	British thermal unit
$^{\circ}$F	degrees Fanrenheit
dia	diameter
FGD	flue gas desulfurization
ft	feet
ft^2	square feet
ft^3	cubic feet
gal	gallon
gpm	gallons per minute
gr	grain
hp	horsepower
hr	hour
in.	inch
k	thousand
kW	kilowatt
kWh	kilowatthour
lb	pound
L/G	liquid-to-gas ratio in gallons per thousand actual cubic feet of gas at outlet conditions
M	million
mi	mile
mo	month
MW	megawatt
ppm	parts per million
psig	pounds per square inch (gauge)
rpm	revolutions per minute
sec	second
sft^3/min	standard cubic feet per minute (60°F)
SS	stainless steel
yr	year

CONVERSION FACTORS

EPA policy is to express all measurements in Agency documents in metric units. Values in this report are given in British units for the convenience of engineers and other scientists accustomed to using the British systems. The following conversion factors may be used to provide metric equivalents.

To convert British		Multiply by	To obtain Metric	
ac	acre	0.405	hectare	ha
Btu	British thermal unit	0.252	kilocalories	kcal
^{o}F	degrees Fahrenheit minus 32	0.5556	degrees Celsius	^{o}C
ft	feet	30.48	centimeters	cm
ft^2	square feet	0.0929	square meters	m^2
ft^3	cubic feet	0.02832	cubic meters	m^3
ft/min	feet per minute	0.508	centimeters per second	cm/s
ft^3/min	cubic feet per minute	0.000472	cubic meters per second	m^3/s
gal	gallons (U.S.)	3.785	liters	L
gpm	gallons per minute	0.06308	liters per second	L/s
gr	grains	0.0648	grams	g
gr/ft^3	grains per cubic foot	2.288	grams per cubic meter	g/m^3
hp	horsepower	0.746	kilowatts	kW
in.	inches	2.54	centimeters	cm
lb	pounds	0.4536	kilograms	kg
lb/ft^3	pounds per cubic foot	16.02	kilograms per cubic meter	kg/m^3
lb/hr	pounds per hour	0.126	grams per second	g/s
psi	pounds per square inch	6895	pascals (newton per square meter)	Pa (N/m^2)
mi	miles	1609	meters	m
rpm	revolutions per minute	0.1047	radians per second	rad/s
sft^3/min	standard cubic feet per minute (60^{o}F)	1.6077	normal cubic meters per hour (0^{o}C)	m^3/hr (0^{o}C)
ton	tons (short)	0.9072	metric tons	tonne
ton/hr	tons per hour	0.252	kilograms per second	kg/s

EXECUTIVE SUMMARY

INTRODUCTION

This study evaluates the economics of limestone and lime wet scrubbing flue gas desulfurization (FGD) processes that incorporate recent technological developments such as forced oxidation, refined absorber design and operating conditions, landfill disposal, and the use of additives to enhance scrubbing efficiency. Most of the processes are based on EPA-performed or EPA-sponsored FGD evaluations, particularly those at the Shawnee test facility. The economic data were obtained using a computer model developed from the Shawnee test facility results.

The standard limestone process, used in TVA economic evaluations since 1967, has evolved from a simple process with pond disposal to a refined design with forced oxidation and landfill disposal. While this represents what is perceived to be a similar general industry evolution, other absorber designs, processes without forced oxidation, pond disposal, and lime absorbent remain important. Recently, the use of additives to enhance SO_2 absorption efficiency has also become a factor in limestone and lime FGD technology.

The primary focus of this study is on absorber type, forced oxidation, and the use of additives. The three types of absorbers [Turbulent Contact Absorber (TCA), spray tower, and venturi - spray tower], the forced oxidation method (single loop, two loop, and bleedstream), and the type of additive (magnesium oxide and adipic acid) are the main subjects. Landfill disposal is the primary disposal method used but a range of pond disposal processes are included for comparison. In this evaluation all processes that do not incorporate forced oxidation have pond disposal. All processes incorporating forced oxidation have landfill disposal, with two exceptions.

Forced oxidation to gypsum has recently evolved into the mainstream of limestone and lime processes. Oxidation to nearly 100% gypsum is achieved and the waste is readily dewatered to 85% or more solids. It is apparently tractable to landfill disposal without stabilization. Use of the gypsum for wallboard and cement manufacture is also a possibility.

The use of magnesium oxide to enhance SO_2 removal efficiency has been investigated for several years, mostly in lime processes. The effect is a result of magnesium and calcium sulfur salt solubility relationships and complex formation that increase dissolved alkalinity. Adipic acid, an organic additive which has recently received considerable attention, acts as a buffer that tends to maintain a liquid pH favorable to SO_2 absorption. It has been extensively evaluated by EPA as a means of improving limestone FGD performance.

The computerized Shawnee lime/limestone scrubbing model used to calculate the design and cost data for this study was developed by Bechtel National, Inc., and TVA to model limestone and lime wet scrubbing. The model consists of subsystems and subroutines that determine FGD operating conditions, design specifications, capital investments, and revenue requirements for specified power unit and coal conditions. The basic processes in the model consist of limestone and lime scrubbing; TCA, spray tower, and venturi absorbers; and pond or landfill disposal. Process options include the three modes of forced oxidation, magnesium addition, and adipic acid addition. Several dozen additional input and output options provide further flexibility in use of the model.

DESIGN AND ECONOMIC PREMISES

The economic evaluations are based on a conceptual design developed from the design premises, engineering data, and equipment costs. All FGD designs are based on SO_2 removal to meet the 1979 new source performance standards (NSPS).

Design Premises

The coals used are three eastern bituminous coals and one western subbituminous coal. The compositions of the coals are shown below. The 3.5% sulfur coal is the base-case coal.

Coal	Sulfur, %	Ash, %	Moisture, %	Heat content, Btu/lb
		(As-Fired Basis)		
Eastern bituminous, 5.0% S	4.80	15.10	4.0	11,700
Eastern bituminous, 3.5% S	3.36	15.14	4.0	11,700
Eastern bituminous, 2.0% S	1.92	15.08	4.0	11,700
Western subbituminous, 0.7% S	0.48	6.30	29.3	8,200

The power plant site is in the north-central region. The base-case power unit is a new, single 500-MW, balanced-draft, horizontally fired, dry-bottom boiler burning pulverized coal. Power unit size case variations consist of similar 200- and 1,000-MW units. The power units have a 30-year life and operate at full load for 5,500 hours a year for a total lifetime operation of 165,000 hours.

Flue gas compositions are based on a total air rate equivalent to 139% of the stoichiometric requirement. It is assumed that 80% of the ash in the coal is emitted as fly ash and 92% of the sulfur is emitted as SO_x (97% SO_2 and 3% SO_3). Particulate matter removal upstream from the FGD system is assumed and the costs for ash collection and disposal are not included in the economic evaluations.

FGD systems are close coupled to the power unit by a plenum into which the power plant ID fans discharge. Separate ID fans are provided for the FGD system. The FGD costs include FGD-related ductwork and associated equipment between the power plant ID fans and the stack plenum. Absorber trains are sized for a maximum of 125 MW of flue gas up to a maximum of 513,000 sft^3/min (60^{o}F). Spare absorber trains for a minimum 25% redundancy and provision for emergency bypass of 50% of the gas that would normally be scrubbed are included. Partial bypass is used for the low-sulfur (0.7% sulfur) coal cases. The disposal site is one mile from the FGD facilities.

Economic Premises

A 3-year construction period, from early 1981 to late 1983, is used. Mid-1982 costs are used for capital investment. Mid-1984 costs are used for annual revenue requirements.

The capital investment estimates consist of direct investment, indirect investment, and other capital investment. Direct investment consists of the installed costs of all process equipment, including provisions for services and miscellaneous expenses, and waste disposal investment. Indirect capital investment covers engineering design and supervision, architect and engineering contractor, construction expense, contractor fees, and contingency. Allowances for startup and modifications and interest during construction are included as other capital investment.

Annual revenue requirements consist of various direct and indirect operating and maintenance costs and capital charges. Both first-year and levelized annual revenue requirements are reported. Levelized capital charges are used in both first-year and levelized annual revenue requirements. Levelized operating and maintenance costs are determined by multiplying by a 30-year levelizing factor of 1.886. This represents an annual inflation rate of 6% and a discount rate of 10%.

SYSTEMS ESTIMATED

The processes evaluated consist primarily of combinations of (1) limestone or lime absorbent; (2) TCA, spray tower, or venturi - spray tower absorber; (3) onsite pond disposal of sludge without forced oxidation or forced oxidation with landfill; and (4) processes without additive, and processes with magnesium oxide or adipic acid additive. The alternatives of increased partial scrubbing (using adipic acid to enhance SO_2 removal efficiency) are also investigated. The cases evaluated are shown in Table S-1. All of the processes are divided into materials handling, feed preparation, gas handling, SO_2 absorption, reheat, oxidation (when used), solids separation (when used), and disposal areas. The materials handling and feed preparation areas consist of equipment to receive, store, and prepare absorbent slurry. Limestone is crushed and wet ball milled. Lime is slaked and slurried. For processes that use additives, additional equipment is included in the preparation area for their addition. The gas handling area consists of a feed plenum that distributes the flue gas to the individual absorber trains, ductwork, and one ID fan per absorber train. Two emergency bypass ducts are included to handle 50% of the scrubbed gas and serve as bypass ducts if some of the flue gas is normally bypassed. The absorption area consists of neoprene-lined carbon steel TCA, spray tower, or venturi - spray tower absorbers, and the absorbent liquid recirculation system. For processes without forced oxidation all three types of absorbers are provided with a single hold tank beneath the absorber.

TABLE S-1. PROCESSES EVALUATED

Case variations	TCA-P	ST-P	V-ST-P	TCA-ILFO-L	ST-ILFO-L	V-ST-VTFO-L	TCA-AA-ILFO-L	ST-AA-ILFO-L	V-ST-AA-VTFO-L	TCA-MgO-BSFO-L	ST-MgO-BSFO-L	V-ST-MgO-VTFO-L	ST-AA-P	ST-MgO-P
New plant														
200 MW														
0.7% S[a]					x									
2.0% S					x									
3.5% S		x			x			x						
5.0% S					x									
500 MW														
0.7% S[a]		x[d]			x[ce]			x[e]			x			
2.0% S		x[d]			x[ce]			x[e]			x			
3.5% S	x[d]	x[d]	x[d]	x[d]	x[bdeg]	x[d]	x	x[ef]	x	x[d]	x[d]	x[d]	x	x
5.0% S		x[d]			x			x			x			
1000 MW														
0.7% S[a]					x									
2.0% S					x									
3.5% S		x			x			x						
5.0% S					x									

PROCESS ABBREVIATIONS:

Scrubbing alternatives

V - Venturi
ST - Spray tower
TCA - Mobile bed absorber
V-ST - Venturi - spray tower

Sludge treatment and disposal alternatives

P - Ponding
L - Landfill
ILFO - In-loop forced oxidation
BSFO - Bleedstream forced oxidation
VTFO - Venturi hold tank forced oxidation

Additive alternatives

AA - Adipic acid
MgO - Magnesium oxide

---

a. Forced oxidation is not applied in these cases, but 95% natural oxidation is assumed.
b. Base case process.
c. Includes cases of full scrubbing and partial scrubbing.
d. Limestone and lime.
e. SO_2 removal efficiencies of NSPS and 95%.
f. Includes a dibasic acid case.
g. Includes a no reheat case.

For processes using the TCA or spray tower with forced oxidation the tank beneath the absorber serves as the oxidation tank and a separate hold tank is provided. For processes using the venturi - spray tower with forced oxidation a separate hold tank for the venturi serves as the oxidation tank. For processes with forced oxidation and MgO additive, bleedstream oxidation is used. All FGD systems are designed for a flue gas temperature of 175°F at the entrance to the stack. Reheat is provided by indirect inline steam reheat with provisions to account for compression by the ID fans and bypassed flue gas. The oxidation area consists of air compressors, the oxidation tank itself, a sparging ring, and an agitator.

In cases without forced oxidation there is no solids separation area. The waste stream (8% solids) is pumped to the pond and pond supernate is returned to the scrubbing area. In cases with forced oxidation the gypsum waste is dewatered with thickeners and rotary vacuum filters to 85% solids. Trucks are used to transport the filter cake to the landfill.

RESULTS

The overall economic results consist of the capital investment in mid-1982 dollars, first-year annual revenue requirements (which include levelized capital charges) in mid-1984 dollars, and levelized (over a 30-year period) annual revenue requirements (which consist of levelized operating and maintenance costs as well as levelized capital charges). Summaries of these costs are shown in Tables S-2, S-3, S-4, and S-5.

Comparison of Limestone Processes With and Without Forced Oxidation

Costs for comparable limestone processes without forced oxidation and with pond disposal and processes with forced oxidation and landfill are shown in Table S-6. The capital investment and levelized annual revenue requirements relationships are shown graphically in Figure S-1 for 500-MW cases. Processes without forced oxidation have higher capital investments than comparable processes with forced oxidation in all cases. The differences increase with increasing coal sulfur content. The higher capital investment is largely a result of disposal area construction costs, which are two to three-and-one-half times higher for ponds than for landfills. In contrast, the additional capital investment for forced oxidation and sludge dewatering is relatively small.

Except for the 5.0% sulfur coal cases, first-year annual revenue requirements are higher for forced-oxidation processes. Although the forced-oxidation processes have lower capital charges in all cases, this advantage is offset at the lower coal sulfur levels by the higher operating labor, maintenance, and overhead costs associated with the waste dewatering and landfill operations. Levelized annual revenue requirements are higher for the forced-oxidation processes in all cases. This is the result of the higher proportion of operating and maintenance costs in the total annual revenue requirements of the forced-oxidation processes.

TABLE S-2. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR LIMESTONE PROCESSES WITHOUT FORCED OXIDATION

							Annual revenue requirements, 1984$			
		Plant size,	Coal sulfur,	Absorber	Capital investment, 1982$		First-year		Levelized	
Case	Absorbent	MW	%	type	M$	$/kW	M$	Mills/kWh	M$	Mills/kWh
1	Limestone	200	3.5	ST	62.7	314	15.8	14.39	21.7	19.72
2	Limestone	500	0.7	ST	73.2	146	18.1	6.56	24.5	8.91
3	Limestone	500	2.0	ST	90.4	181	23.3	8.46	32.1	11.68
4	Limestone	500	3.5	ST	112.6	225	28.8	10.47	39.6	14.42
5	Limestone	500	5.0	ST	123.3	247	31.5	11.44	43.3	15.73
6	Limestone	500	3.5	TCA	107.8	216	27.6	10.03	38.0	13.81
7	Limestone	500	3.5	VST	121.7	243	31.3	11.37	43.1	15.68
8	Limestone	1,000	3.5	ST	192.1	192	48.9	8.89	67.2	12.23

Note: No additives, pond disposal in all cases.

TABLE S-3. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR LIMESTONE PROCESSES WITH FORCED OXIDATION

Case	Absorbent	Plant size, MW	Coal sulfur, %	Absorber type	Capital investment, 1982$		Annual revenue requirements, 1984$			
							First-year		Levelized	
					M$	$/kW	M$	Mills/kWh	M$	Mills/kWh
9	Limestone	200	0.7	ST	44.4	222	12.3	11.15	17.3	15.76
10	Limestone	200	2.0	ST	52.1	260	14.7	13.40	21.0	19.10
11	Limestone	200	3.5	ST	59.1	295	16.7	15.21	23.9	21.69
12	Limestone	200	5.0	ST	62.6	313	17.8	16.17	25.4	23.08
13	Limestone	500	0.7	ST	70.4	141	18.8	6.84	26.3	9.56
14	Limestone	500	2.0	ST	86.6	173	24.2	8.80	34.4	12.49
15	Limestone	500	3.5	ST	103.1	206	29.1	10.58	41.5	15.08
16	Limestone	500	5.0	ST	109.2	218	31.2	11.36	44.7	16.25
17	Limestone	500	3.5	TCA	98.5	197	27.8	10.09	39.5	14.37
18	Limestone	500	3.5	VST	109.2	218	30.8	11.20	43.9	15.96
19	Limestone[a]	500	3.5	ST	94.4	189	25.8	9.40	36.4	13.25
20	Limestone	1,000	0.7	ST	109.2	109	28.3	5.15	39.2	7.13
21	Limestone	1,000	2.0	ST	151.7	152	41.1	7.47	57.7	10.50
22	Limestone	1,000	3.5	ST	177.2	177	49.3	8.97	69.9	12.71
23	Limestone	1,000	5.0	ST	187.0	187	53.0	9.63	75.5	13.73

Note: No additives, landfill disposal in all cases.

[a] No reheat.

TABLE S-4. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR LIMESTONE PROCESSES WITH FORCED OXIDATION AND ADDITIVES

Case	Absorbent	Plant size, MW	Coal sulfur, %	Absorber type	Capital investment, 1982$		Annual revenue requirements, 1984$			
							First-year		Levelized	
					M$	$/kW	M$	Mills/kWh	M$	Mills/kWh
Adipic Acid										
24	Limestone	200	3.5	ST	54.4	272	15.3	13.95	21.9	19.86
25	Limestone	500	0.7	ST	61.4	123	16.4	5.97	23.0	8.36
26	Limestone	500	2.0	ST	79.1	158	21.6	7.86	30.4	11.07
27	Limestone	500	3.5	ST	92.8	186	25.8	9.38	36.6	13.31
28	Limestone	500	5.0	ST	97.9	196	27.7	10.07	39.5	14.36
29	Limestone	500	3.5	TCA	90.4	181	25.4	9.23	36.1	13.12
30	Limestone	500	3.5	VST	96.3	193	27.0	9.81	38.3	13.93
31	Limestone[a]	500	3.5	ST	100.7	201	26.8	9.73	37.3	13.58
32	Limestone[b]	500	3.5	ST	92.8	186	25.7	9.35	36.4	13.25
33	Limestone	1,000	3.5	ST	155.5	155	42.8	7.78	60.4	10.99
Magnesium Oxide										
34	Limestone	500	0.7	ST	69.1	138	18.4	6.70	25.8	9.37
35	Limestone	500	2.0	ST	86.6	173	23.8	8.66	33.6	12.23
36	Limestone	500	3.5	ST	102.5	205	28.8	10.49	41.1	14.93
37	Limestone	500	5.0	ST	107.9	216	30.7	11.18	43.9	15.98
38	Limestone	500	3.5	TCA	98.2	196	27.7	10.09	39.5	14.38
39	Limestone	500	3.5	VST	113.1	226	31.6	11.50	44.9	16.33
40	Limestone[a]	500	3.5	ST	111.2	222	28.5	10.36	39.2	14.27

Note: Landfill disposal except as noted.

a. Pond disposal.

b. Dibasic acid.

TABLE S-5. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR LIME PROCESSES

Case	Absorbent	Plant size, MW	Coal sulfur, %	Absorber type	Capital investment, 1982$		Annual revenue requirements, 1984$			
							First-year		Levelized	
					M$	$/kW	M$	Mills/kWh	M$	Mills/kWh
Without Forced Oxidation (Pond Disposal)										
41	Lime	500	0.7	ST	60.7	121	15.3	5.56	20.9	7.62
42	Lime	500	2.0	ST	79.3	159	21.8	7.92	30.7	11.18
43	Lime	500	3.5	ST	96.2	192	28.0	10.19	40.3	14.66
44	Lime	500	5.0	ST	105.3	211	31.6	11.51	46.0	16.71
45	Lime	500	3.5	TCA	94.6	189	27.7	10.09	40.0	14.55
46	Lime	500	3.5	VST	103.2	206	29.9	10.88	43.0	15.63
With Forced Oxidation (Landfill Disposal)										
47	Lime	500	3.5	ST	88.9	178	28.5	10.35	42.1	15.31
48	Lime	500	3.5	TCA	86.2	172	27.9	10.16	41.5	15.07
49	Lime	500	3.5	VST	92.7	185	29.6	10.77	43.8	15.92
With Forced Oxidation and Magnesium Oxide (Landfill Disposal)										
50	Lime	500	3.5	ST	89.6	179	28.6	10.40	42.3	15.37
51	Lime	500	3.5	TCA	90.4	181	28.7	10.45	42.4	15.43
52	Lime	500	3.5	VST	96.8	194	30.3	11.01	44.5	16.18

TABLE S-6. COST COMPARISON OF PROCESSES
WITH AND WITHOUT FORCED OXIDATION

Process[a]	Capital investment, 1982 $/kW	Annual revenue requirements, 1984 mills/kWh First-year	Levelized
0.7% Sulfur Coal			
No forced oxidation, pond	146.5	6.56	8.91
No forced oxidation, landfill	140.7	6.84	9.56
2.0% Sulfur Coal			
No forced oxidation, pond	180.8	8.46	11.68
Forced oxidation, landfill	173.1	8.80	12.49
3.5% Sulfur Coal			
No forced oxidation, pond	225.1	10.47	14.42
Forced oxidation, landfill	206.1	10.58	15.08
5.0% Sulfur Coal			
No forced oxidation, pond	246.7	11.44	15.73
Forced oxidation, landfill	218.3	11.36	16.25

[a] 500-MW, limestone spray tower processes, natural oxidation to 95% $CaSO_4 \cdot 2H_2O$ for 0.7% sulfur coal cases, in-loop forced oxidation to 95% $CaSO_4 \cdot 2H_2O$ for all other cases.

Additives

Costs of comparable processes with and without additives are shown in Table S-7.

Magnesium Oxide--

The use of magnesium oxide as an additive in the limestone process has no significant effect on capital investment. The equipment needed to handle the relatively small volumes of additive does not materially affect equipment costs. By improving SO_2 removal efficiency and absorbent utilization the use of magnesium oxide slightly decreases gas handling, SO_2 absorption, and

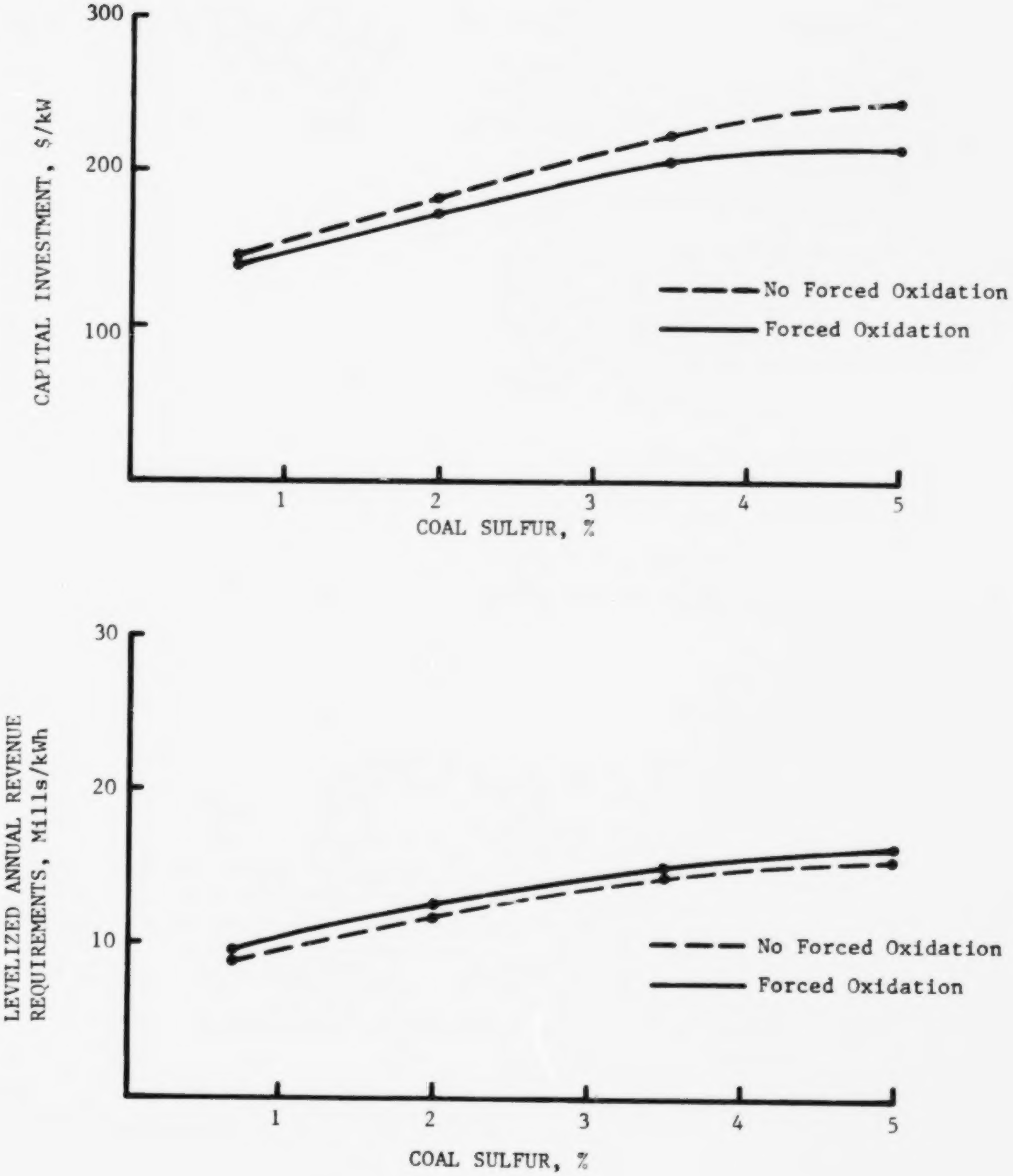


Figure S-1. Cost comparison of 500-MW limestone processes with and without forced oxidation at different coal sulfur contents.

TABLE S-7. COST COMPARISON OF PROCESSES WITH AND WITHOUT ADDITIVES

	Capital investment, 1982 $/kW	Annual revenue requirements, 1984 mills/kWh	
		First-year	Levelized
Landfill Disposal[a]			
0.7% sulfur coal			
No additive	140.7	6.84	9.56
Adipic acid	122.8	5.97	8.36
MgO	138.2	6.70	9.37
2.0% sulfur coal			
No additive	173.1	8.80	12.49
Adipic acid	158.2	7.86	11.07
MgO	171.8	8.72	12.39
3.5% sulfur coal			
No additive	206.1	10.58	15.08
Adipic acid	185.5	9.38	13.31
MgO	205.0	10.49	14.93
5.0% sulfur coal			
No additive	218.3	11.36	16.25
Adipic acid	195.9	10.07	14.36
MgO	215.8	11.18	15.98
Pond Disposal[b]			
3.5% sulfur coal			
No additive	225.1	10.47	14.42
Adipic acid	201.4	9.73	13.58
MgO	222.4	10.36	14.27

a. 500-MW, limestone spray tower processes, natural oxidation for all 0.7% sulfur coal cases, in-loop forced oxidation for no additives and adipic acid cases, bleedstream forced oxidation for all MgO processes.

b. 500-MW, limestone spray tower processes without forced oxidation.

disposal costs. The effect is thus more pronounced at higher coal sulfur contents. For the 3.5% sulfur coal case the reduction in capital investment is less than 1%, however.

The effect of magnesium oxide on first-year annual revenue requirements of forced-oxidation processes is similar to the effect on capital investment. The cost of the additive is minimal (less than 1% of total annual revenue requirements). The improved SO_2 removal efficiency and absorbent utilization, expressed primarily as reduced absorbent and electricity costs, more than offset the cost for magnesium oxide. For the 3.5% sulfur coal case the cost reduction is less than 1%. In terms of levelized annual revenue requirements the results are essentially the same, i.e., the use of magnesium oxide decreases the overall annual revenue requirements but the decrease is relatively insignificant.

Adipic Acid--

In contrast to the results with magnesium oxide addition, the use of adipic acid appears to be an economically attractive additive for all of the applications considered in this study. Capital investment reductions range from nearly 13% for the 0.7% sulfur western subbituminous coal case to about 10% for the high-sulfur coal cases. As in the magnesium oxide addition cases, the additional equipment needed to handle the small quantities of adipic acid does not have an appreciable effect on equipment costs. However, by improving SO_2 removal efficiency and absorbent utilization the use of adipic acid decreases the capital investments for the feed preparation, gas handling, SO_2 absorption, solids separation, and disposal area costs. The primary reasons for the more dramatic reductions in costs are the assumptions that with adipic acid the limestone stoichiometry can be reduced by 24% (1.07 versus 1.40) and that 95% SO_2 removal can be achieved in the absorber and thus partial flue gas bypass can be used to reduce both the reheat requirements and the size of the absorbers. (It should be noted that vendors are quoting stoichiometries as low as 1.10 on bids and this would obviously result in a lower cost advantage for the adipic acid-enhanced process.)

The effect of adipic acid addition on both the first-year and the levelized annual revenue requirements for the forced-oxidation processes is similar to its effect on capital investment: It decreases costs 11% to 13% depending on the coal sulfur level. Most of this cost reduction is due to a saving in capital charges. Other reductions such as those for electricity, absorbent, maintenance, and overheads also contribute to a significant degree. In all forced-oxidation cases the cost of adipic acid is minimal (less than 1% of annual revenue requirements for all except the 5.0% sulfur coal case). The improved SO_2 removal efficiency and absorbent utilization, expressed primarily as reduced absorbent and electricity costs, more than offset the cost for adipic acid for all cases.

In processes without forced oxidation the annual revenue requirements of processes using additives are lower than those of the same process without additives, but the difference is less because of the additional additive lost in the low-solids wastes. The cost of the additive in these cases is approximately the same as the absorbent cost. Except for the additional additive costs, the other cost reductions remain unchanged.

Although cost differences of these magnitudes would not normally be regarded as highly significant, the nature of the comparisons allows more decisive conclusions in this case. The processes are very similar except for the use of additives. Excluding extreme variations in additive cost (relative to other cost variations) the improvements in absorbent utilization and SO_2 removal efficiency indicate that additives provide a real cost advantage for high-sulfur coal landfill applications.

Power Unit Size and Sulfur Content

Power unit size comparisons show the usual effect of economies of scale in terms of unit costs ($/kW, mills/kWh). Comparable processes for 500-MW cases are about 33% less expensive in $/kW and mills/kWh than 200-MW cases. Similarly, 1,000-MW cases are about 13% less expensive than 500-MW cases.

Absorber Type

Cost comparisons for limestone processes with different types of absorbers are shown in Table S-8 for limestone processes at the 500-MW, 3.5% sulfur coal conditions. The capital investment relationships of the processes with TCA's, spray towers, and venturi - spray towers remain the same whether forced oxidation or additives are used. In all cases the TCA process has the lowest capital investment and the venturi - spray tower has the highest.

The major capital investment advantage of the TCA process lies in lower costs for SO_2 absorption because of the smaller absorber and lower L/G. Processes using the larger, lower pressure drop spray tower have higher capital investments for SO_2 absorption than the TCA processes, which is only partially offset by the lower capital investment for fans. The processes using the venturi - spray tower absorbers have a higher limestone utilization rate, and thus slightly lower capital investments for feed preparation, solids disposal (dewatering), and disposal area costs. This is more than offset by the higher capital investments for the gas handling and SO_2 absorption areas. The same cost relationships also prevail in annual revenue requirements (both first-year and levelized) largely because of costs based on capital investment such as maintenance and capital charges. In costs other than maintenance and capital charges there is little difference in annual revenue requirements between the processes with TCA's, spray towers, or venturi - spray towers.

Spray Tower Absorber Design Conditions

Since the superficial gas velocity through the absorber, the pH of the slurry, and the SO_2 removal efficiency are normally determined by the technology or the performance requirements, the absorber liquid to gas (L/G) ratio and stoichiometry are the only remaining design conditions that can be optimized. Thus it becomes an economic tradeoff between the increased costs associated with larger slurry recirculation pumps (L/G) and the costs for the limestone preparation equipment and the waste disposal area (stoichiometry).

TABLE S-8. COST COMPARISON BY ABSORBER TYPE

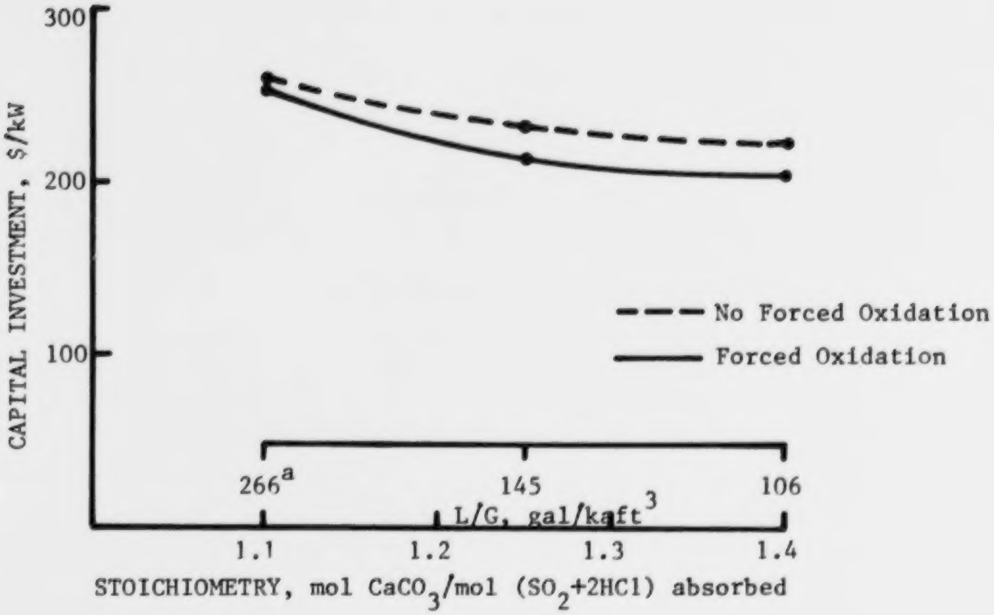
Process[a]	Capital investment, 1982 $/kW	Annual revenue requirements, 1984 mills/kWh First-year	Annual revenue requirements, 1984 mills/kWh Levelized
Without Forced Oxidation (Pond Disposal)			
Spray tower	225.1	10.47	14.42
TCA	215.6	10.03	13.81
Venturi - spray tower	243.4	11.37	15.68
With Forced Oxidation (Landfill Disposal)			
Spray tower	206.1	10.58	15.08
TCA	197.0	10.09	14.37
Venturi - spray tower	218.4	11.20	15.96
With Forced Oxidation and Adipic Acid (Landfill Disposal)			
Spray tower	185.5	9.38	13.31
TCA	180.9	9.23	13.12
Venturi - spray tower	192.7	9.81	13.93

a. 500 MW.

As shown in Figure S-2, increasing the limestone stoichiometry and decreasing the absorber L/G is the more economical option. Increasing the stoichiometry from 1.1 to 1.4 mol $CaCO_3$/mol (SO_2 + 2HCl) absorbed results in a decrease in capital investment of 18% for the forced-oxidation case and nearly 14% for the case without forced oxidation. Levelized annual revenue requirements decrease 18% for the forced-oxidation case and more than 16% for the case without forced oxidation.

Partial Scrubbing Versus Full Scrubbing

For lower sulfur coal applications, where relatively modest SO_2 removal efficiencies are required, two alternative designs are possible: full scrubbing at the required removal efficiency or scrubbing part of the flue gas at a high removal efficiency (i.e., 90%) and bypassing the rest of the flue gas around the absorber to reduce or eliminate the need for stack gas reheat. Figure S-3 shows the costs of full scrubbing and partial scrubbing for a range of coal sulfur levels. These costs are based on a 500-MW limestone process



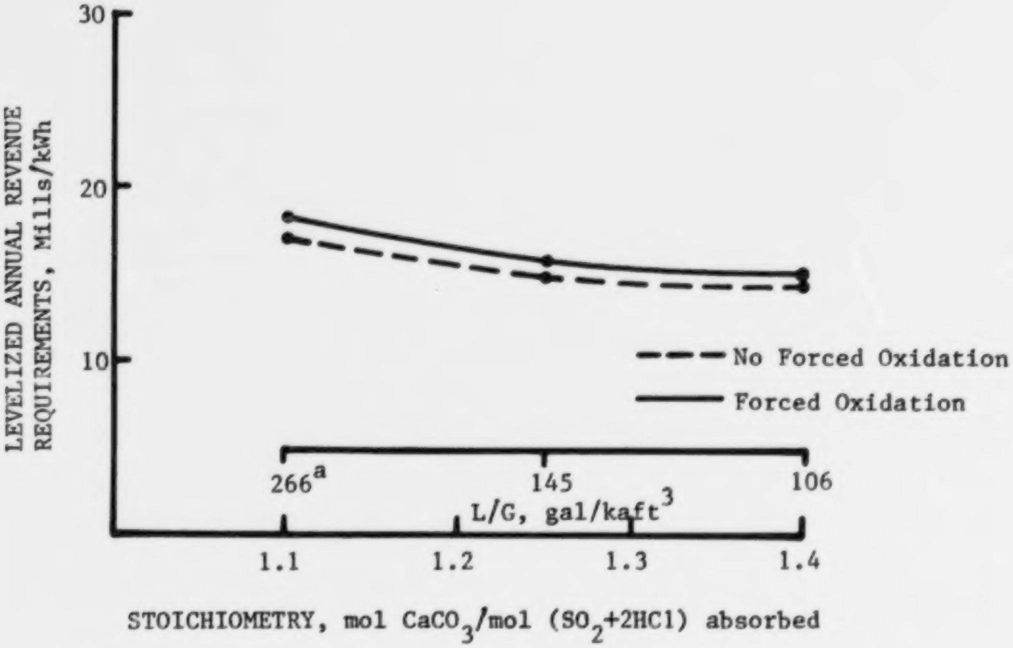


Figure S-2. Cost comparison of limestone processes with and without forced oxidation at different absorber stoichiometries.

a. Extrapolation

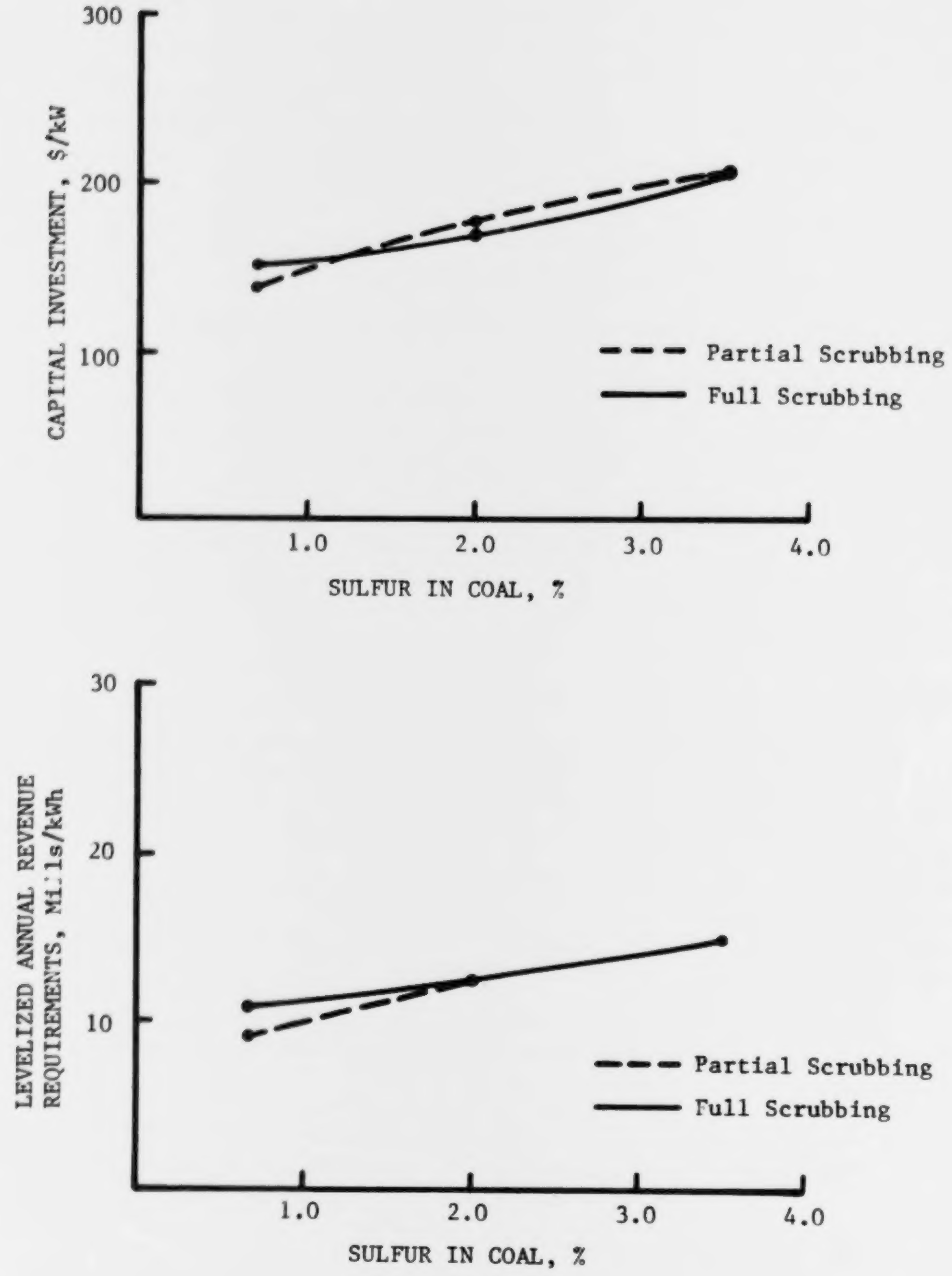


Figure S-3. Cost comparison of 500-MW limestone forced-oxidation processes with partial or full scrubbing at different coal sulfur contents.

with a spray tower absorber, forced oxidation, and landfill disposal. The partial scrubbing cases are based on 90% SO_2 removal in the absorber. For coal sulfur levels below 2% it is more economical to partially scrub and bypass the remaining flue gas. In terms of capital investment it is less expensive to partially scrub the flue gas from coals with sulfur contents below about 1.5%. For the 0.7% sulfur western subbituminous coal, partial scrubbing results in a saving of about $10/kW in capital investment (a 6.5% reduction) and about 1.2 mills/kWh in levelized annual revenue requirements (a 11.2% reduction). For the 2.0% sulfur coal case, full scrubbing results in a $4.0/kW saving in capital investment over the partial scrubbing case but the levelized annual revenue requirements are essentially the same.

Full-Scrubbing Processes at 85%, 90%, and 95% SO_2 Removal

In addition to lowering the limestone stoichiometry, another advantage of using adipic acid in a limestone scrubbing process is shown in Figure S-4. The economics of a forced-oxidation process with and without adipic acid are shown as a function of the required SO_2 removal efficiency. As the required SO_2 removal efficiency increases from 85% to 95%, the economic advantage of the adipic acid-enhanced process increases. At 85% removal the adipic acid-enhanced process is about 9% lower in capital investment and nearly 12% lower in levelized annual revenue requirements. At 95% removal the corresponding reductions are more than 15% for both capital investment and levelized annual revenue requirements.

Reheat

Since inline steam reheating involves significant expenditures in both capital investment and annual revenue requirements and the benefits of reheating cannot be determined easily, some utilities have recently begun specifying FGD systems without reheat. Table S-9 compares the economics of two limestone forced-oxidation processes, one with reheat and the other without. It should be noted that this comparison represents the most optimistic assumptions for the no reheat case. In actual practice one might expect maintenance costs for the no reheat case to be at least equal to, if not higher than, those for the reheat case.

The capital investment for the no reheat case is more than 8% lower ($18/kW) than the reheat case. In first-year and levelized annual revenue requirements, the no reheat case is about 11% and 12% lower, respectively. These lower costs are due partly to the elimination of steam consumption and partly to the reduction in maintenance, overheads, and capital charges (these last reductions are due to the lower capital investment for the no reheat case).

Limestone and Lime Processes

Limestone and lime process costs by process area are compared for different coal sulfur contents and by process type in Table S-10. The processes using lime maintain a substantial advantage in capital investment for all conditions. The advantage decreases slightly with increasing coal

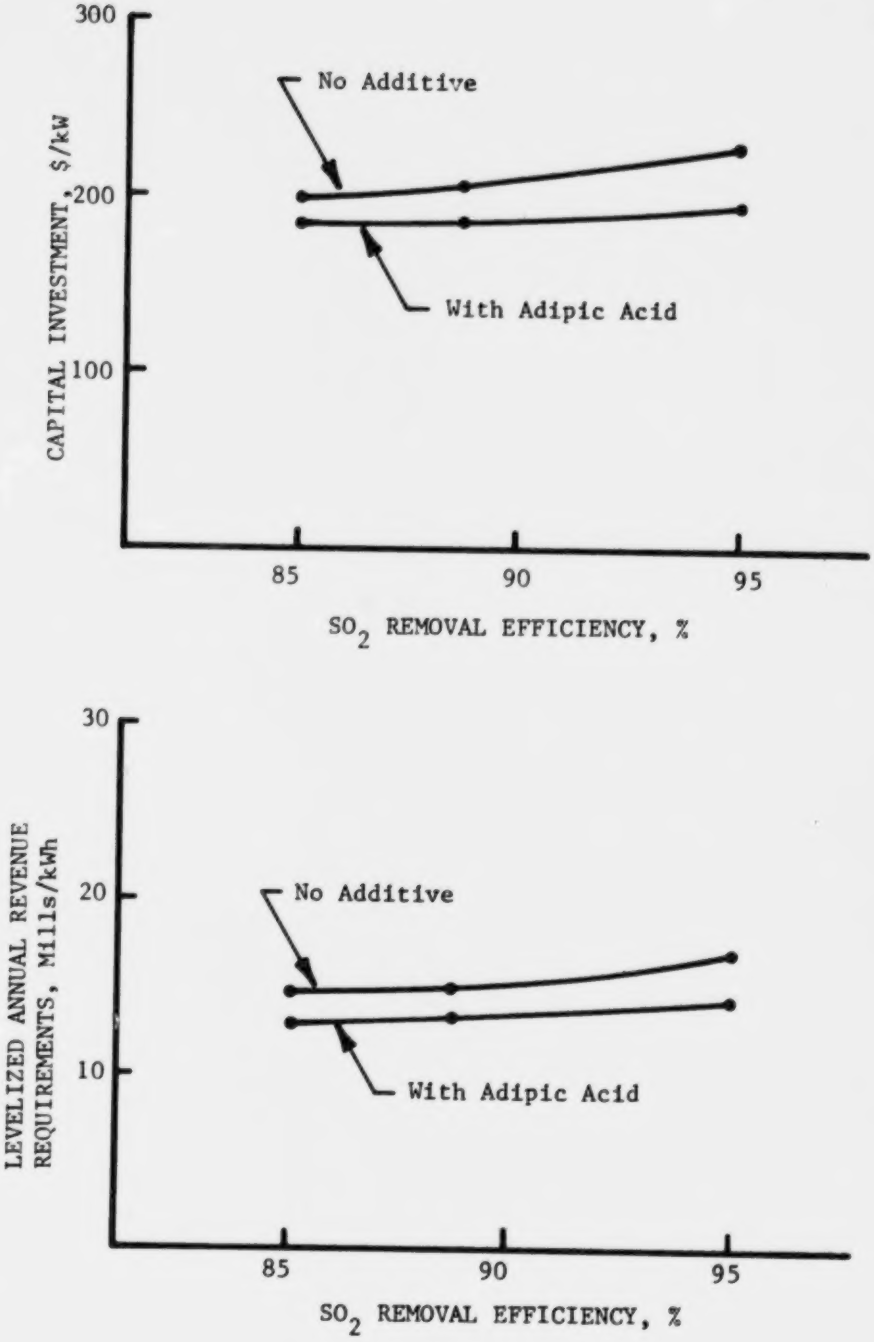


Figure S-4. Cost comparison of 500-MW limestone forced-oxidation processes with and without adipic acid at different SO_2 removal efficiencies

TABLE S-9. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR LIMESTONE PROCESSES WITH AND WITHOUT REHEAT

	Capital investment, 1982 $/kW											
Process[a]	Materials handling	Feed preparation	Gas handling	SO_2 absorption	Oxidation	Reheat	Solids separation	Services, etc.	Disposal area	Total direct	Total indirect	Total
With reheat	5.1	9.4	22.6	40.6	5.3	7.3	7.4	5.8	9.8	113.3	92.8	206.1
Without reheat	5.1	9.4	22.2	39.4	5.3	-	7.4	5.3	9.8	103.9	85.0	188.9

	First-year annual revenue requirements, 1984 mills/kWh										Levelized annual revenue requirements, 1984 mills/kWh
	Absorbent	Labor	Steam	Water	Electricity	Diesel fuel	Maintenance[b]	Overheads	Capital charges	Total	
With reheat	0.44	0.46	0.49	0.01	0.78	0.06	1.60	1.23	5.51	10.58	15.08
Without reheat	0.44	0.46	-	0.01	0.76	0.06	1.46	1.16	5.05	9.40	13.25

a. 500-MW, 3.5% sulfur, limestone spray tower process with forced oxidation and landfill.
b. Includes analysis.

sulfur content (the lime process is 17% lower for the 0.7% sulfur coal case and nearly 15% lower for the 5% sulfur coal case). At the same coal sulfur content the capital investment advantage of the lime processes with variations such as forced oxidation and additives remains relatively constant at about 15% lower than the corresponding limestone processes.

TABLE S-10. COST COMPARISON OF LIMESTONE AND LIME PROCESSES

Process	Capital investment, 1982 $/kW	Annual revenue requirements, 1984 mills/kWh First-year	Annual revenue requirements, 1984 mills/kWh Levelized
Spray Tower[a]			
0.7% sulfur coal			
Limestone	146.5	6.56	8.91
Lime	121.5	5.56	7.62
2.0% sulfur coal			
Limestone	183.4	8.46	11.61
Lime	158.6	7.92	11.81
3.5% sulfur coal			
Limestone	225.1	10.47	14.42
Lime	192.3	10.19	14.66
5.0% sulfur coal			
Limestone	246.7	11.44	15.73
Lime	210.6	11.51	16.71
Venturi - Spray Tower[b]			
Without forced oxidation (pond)			
Limestone	243.4	11.37	15.68
Lime	206.3	10.88	15.63
Forced oxidation (landfill)			
Limestone	218.4	11.20	15.96
Lime	185.4	10.78	15.92
With forced oxidation and MgO[c] (landfill)			
Limestone	226.1	11.50	16.33
Lime	193.5	11.01	16.18

a. 500-MW, without forced oxidation, pond disposal.
b. 500-MW, 3.5% sulfur coal.
c. Bleedstream forced oxidation.

In annual revenue requirements, coal sulfur content has an important effect on the cost relationships of the limestone and lime processes. The lime process has lower first-year annual revenue requirements for all but the

5.0% sulfur coal case. This is due to increasing absorbent costs as the coal sulfur content increases. The lime process, however, has lower levelized annual revenue requirements only for the 0.7% sulfur coal case. The different relationship in first-year and levelized annual revenue requirements occurs because levelizing increases operating and maintenance costs, which are a significantly lower portion of annual revenue requirements for the limestone process. The same cost relationships hold true for the capital investments and first-year annual revenue requirements of forced-oxidation processes, venturi - spray tower processes, and processes with MgO additive. In these cases, however, the limestone process also has higher levelized annual revenue requirements. This is a result of the greater complexity of the processes, which reduces the importance of the lower operating and maintenance costs of the limestone processes.

Cost Sensitivities

The largest components of the annual revenue requirements of limestone processes are capital charges, followed by maintenance, overheads, electricity, steam, absorbent, and labor. There is relatively little difference in the cost distribution of the annual revenue requirements for the various limestone processes. The greatest differences in cost distribution are between the processes without forced oxidation and pond disposal and processes with forced oxidation and landfill disposal. The processes without forced oxidation have relatively lower labor costs and higher capital charges because of the differences between ponding and landfill construction costs. Costs related to forced oxidation and the use of additives are too small a percentage of total annual revenue requirements to affect the relative economics of the processes even if they were to vary radically. In the case of lime processes the absorbent cost is a major factor, exceeding all other cost elements except capital charges. In other component costs, however, the lime processes are similar to limestone processes.

Changes in the Inflation and Discount Rate--

Changes in the inflation and discount rates from 4% inflation and 7% discount rate to 8% inflation and a 13% discount rate (compared with the 6% inflation and 10% discount rates used in this study) have a moderate effect on capital investment and an appreciable effect on annual revenue requirements. The capital investment increases about 8%, the first-year annual revenue requirements increase about 30%, and the levelized annual revenue requirements increase about 45% between the lower and higher rates. The effects are shown in Figure S-5 for the limestone process with forced oxidation and landfill disposal.

Changes in Labor and Maintenance Costs--

The sensitivity of process costs to changes in labor and maintenance costs is shown in Figure S-6 for the same process. Changes in labor cost have very little effect because of the small fraction of annual revenue requirements that constitute labor costs. The effect of changes in maintenance costs is greater, reflecting the 15% fraction of annual revenue requirements

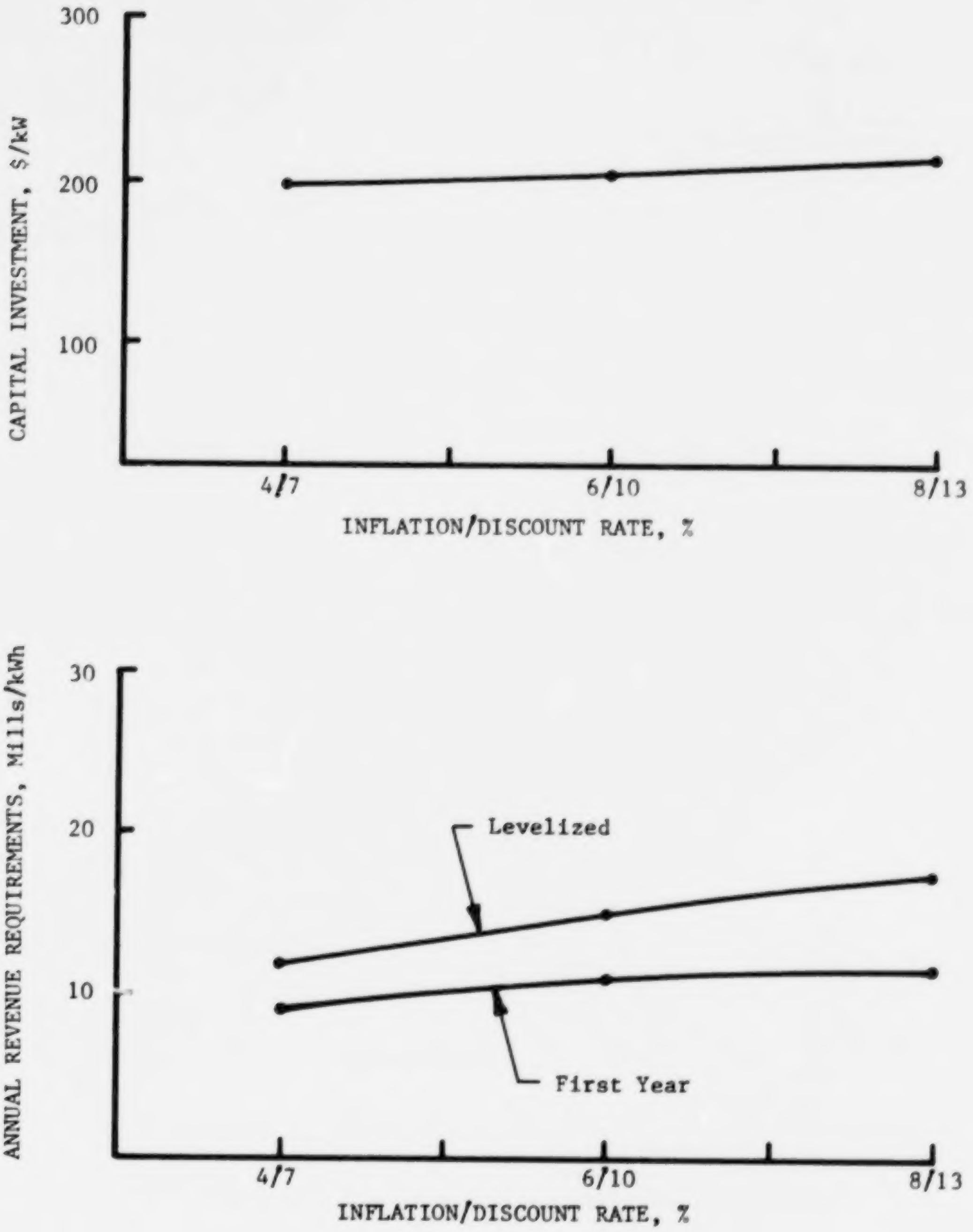


Figure S-5. Sensitivity of the limestone forced-oxidation process costs to changes in inflation and discount rate.

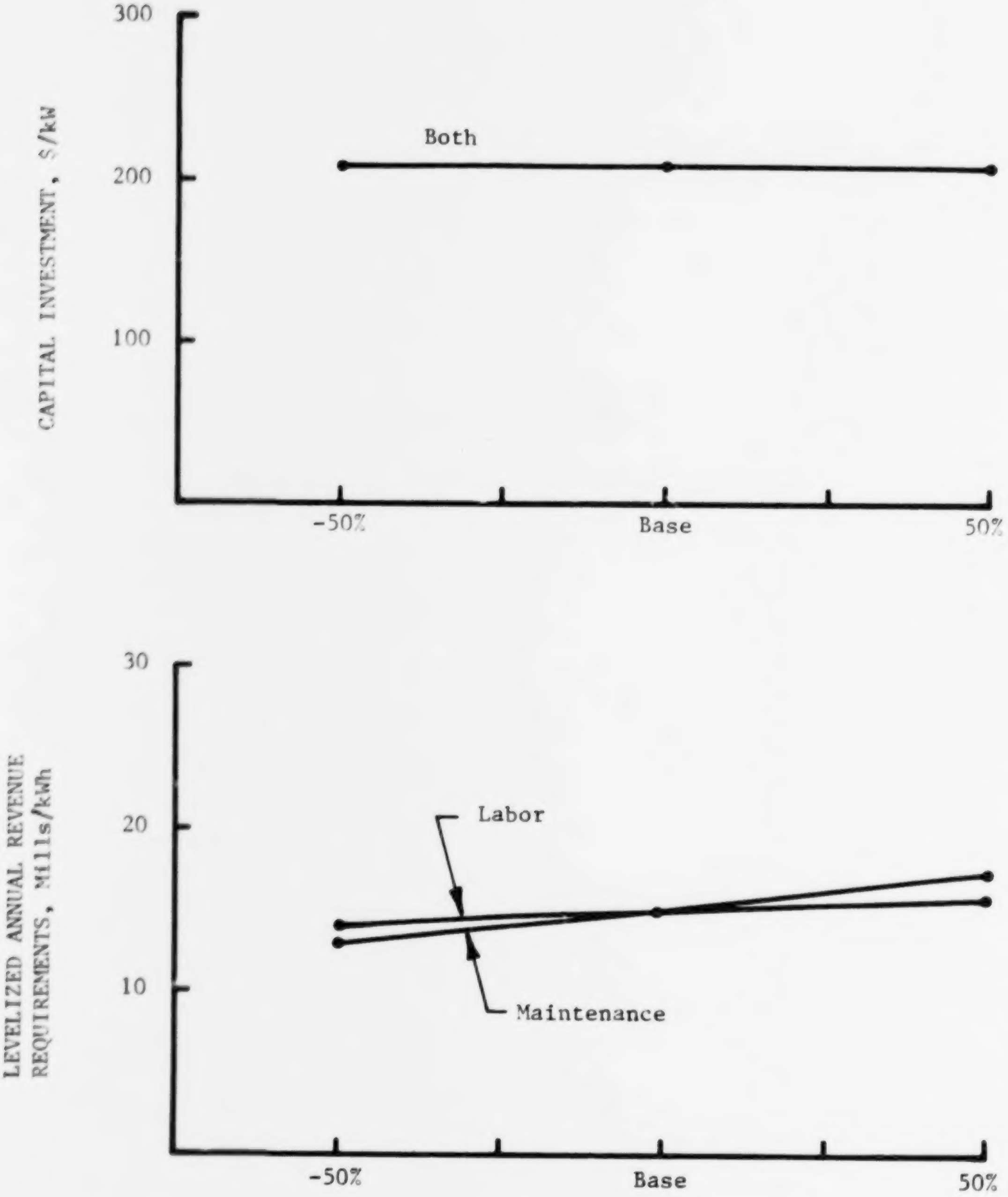


Figure S-6. Sensitivity of limestone forced-oxidation process costs to changes in labor and maintenance costs.

composed of maintenance costs. For example, a change of about 15% in maintenance costs from those used in this study eliminates the difference in annual revenue requirements between the TCA and spray tower processes.

Changes in Lime Costs--

In contrast to the limestone processes, absorbent cost is an appreciable element of the annual revenue requirements for the lime process. This increased sensitivity of the lime process to absorbent costs is shown in Figure S-7. The effect of a 50% increase in the cost of lime on the capital investment is relatively insignificant; however, the first-year and levelized annual revenue requirements increase 8% and 21% respectively.

Use of Dibasic Acid Instead of Adipic Acid--

Questions have been raised about the high unit cost and high degradation rate of adipic acid. One alternative is to use dibasic acid (DBA), rather than adipic acid, because of its lower unit cost. Table S-11 compares the relative process economics for two limestone forced-oxidation processes, one enhanced with adipic acid and the other with DBA. The DBA-enhanced process shows no advantage in terms of capital investment and only a slight decrease (less than 0.5%) in both first-year and levelized annual revenue requirements.

TABLE S-11. COST COMPARISON OF ADIPIC ACID

AND DBA ENHANCED LIMESTONE PROCESSES

	Capital investment, 1982$		Annual revenue requirements, 1984$			
			First-year		Levelized	
Process[a]	M$	$/kW	M$	Mills/kWh	M$	Mills/kWh
Limestone with adipic acid	92.8	186	25.81	9.38	36.59	13.31
Limestone with dibasic acid	92.8	186	25.73	9.35	36.44	13.25

a. 500 MW, 3.5% sulfur coal, spray tower, forced oxidation, landfill.

Changes in Adipic Acid Cost and Degradation Rate--

The sensitivity of process costs to changes in the unit cost for adipic acid is shown in Figure S-8. The capital investment and levelized annual revenue requirements are essentially constant over a 50% cost change. The sensitivity of process costs to changes in the adipic acid degradation rate is shown in Figure S-9. The capital investment and levelized annual revenue requirements are essentially constant over a 200% change in degradation rate. Thus, although adipic acid is an expensive raw material, the consumption rate is low and changes in the cost or consumption rate have a negligible effect on total costs.

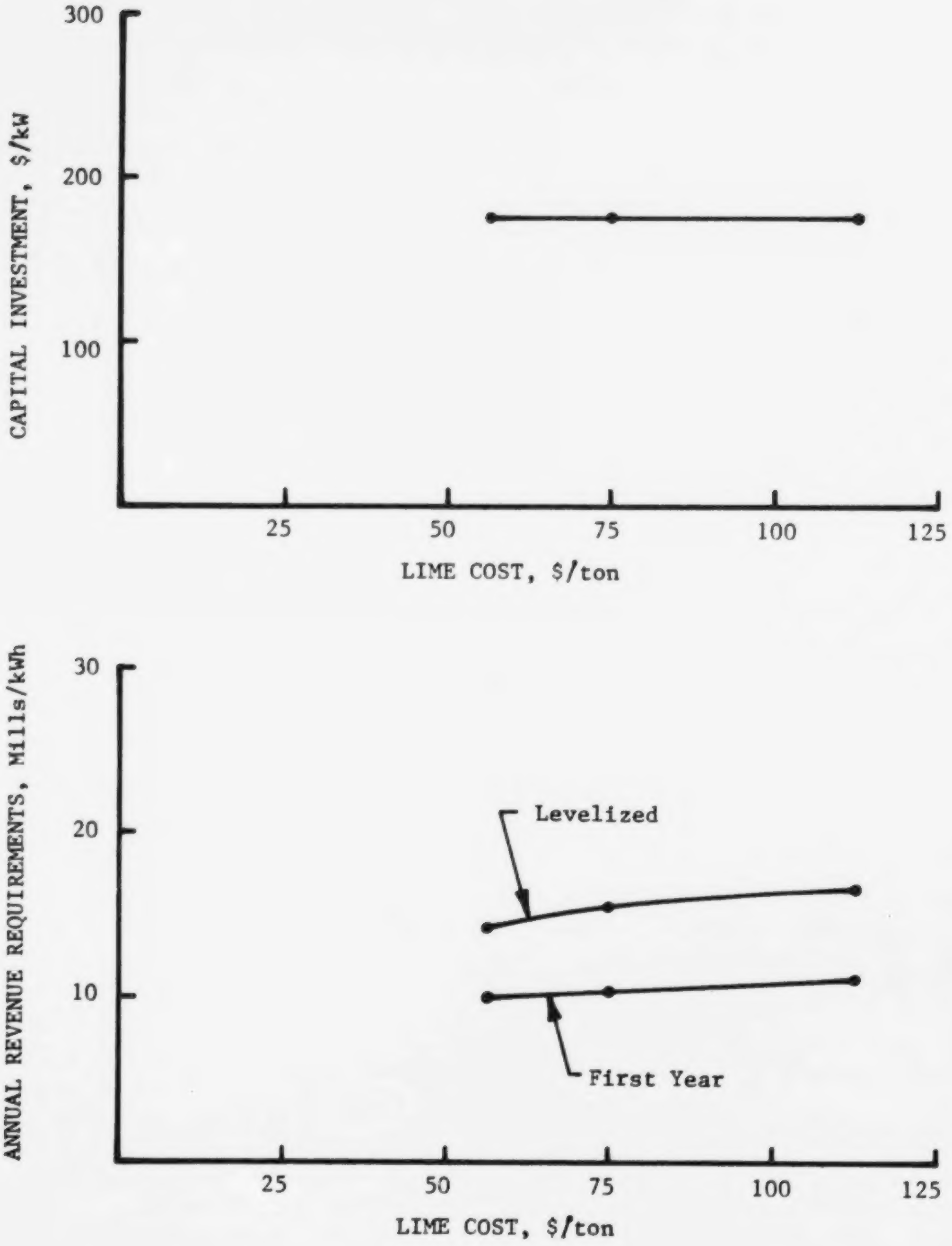


Figure S-7. Sensitivity of lime process costs to the cost of lime.

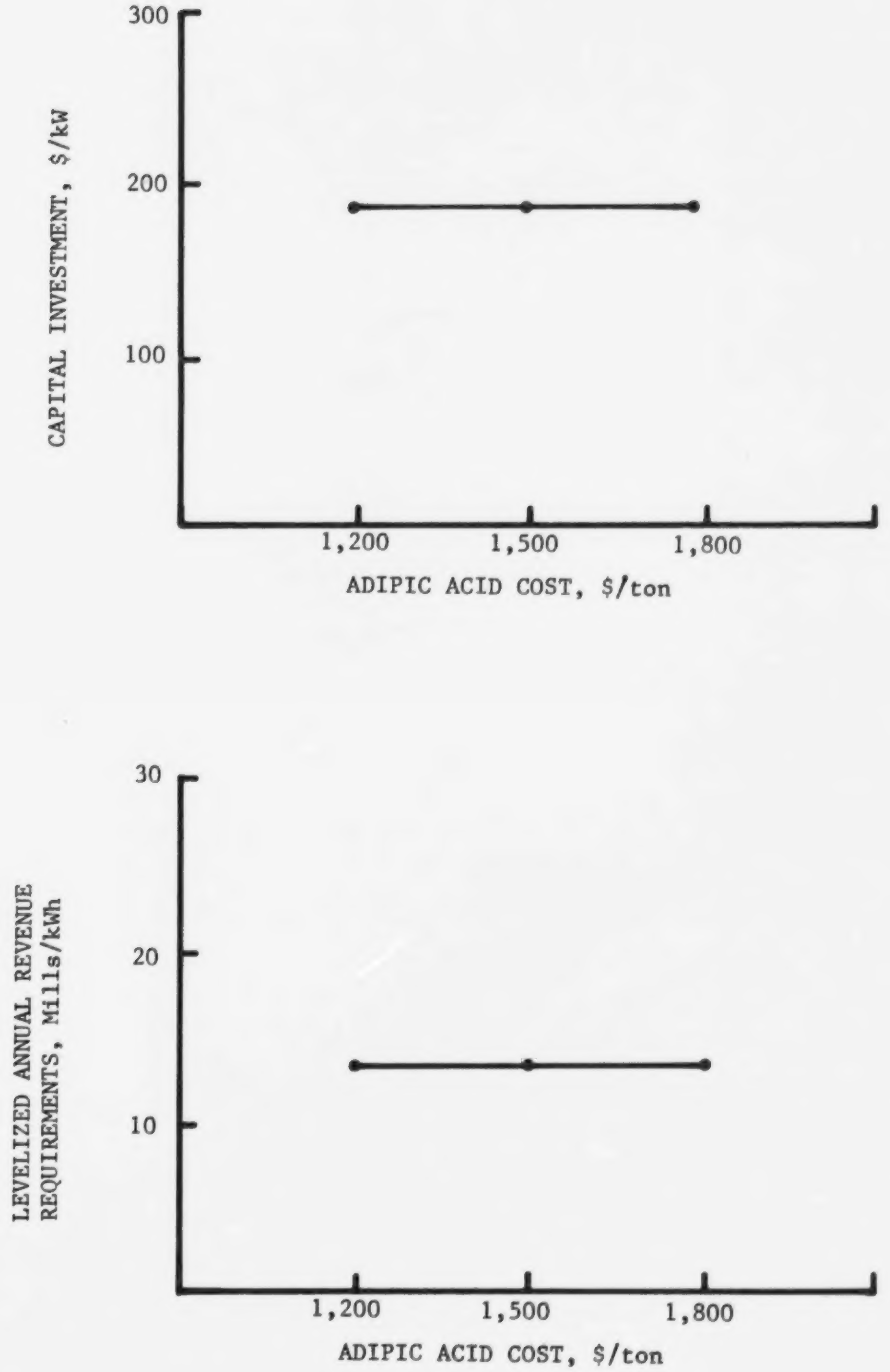


Figure S-8. Sensitivity of limestone adipic acid forced-oxidation process costs to the cost of adipic acid.

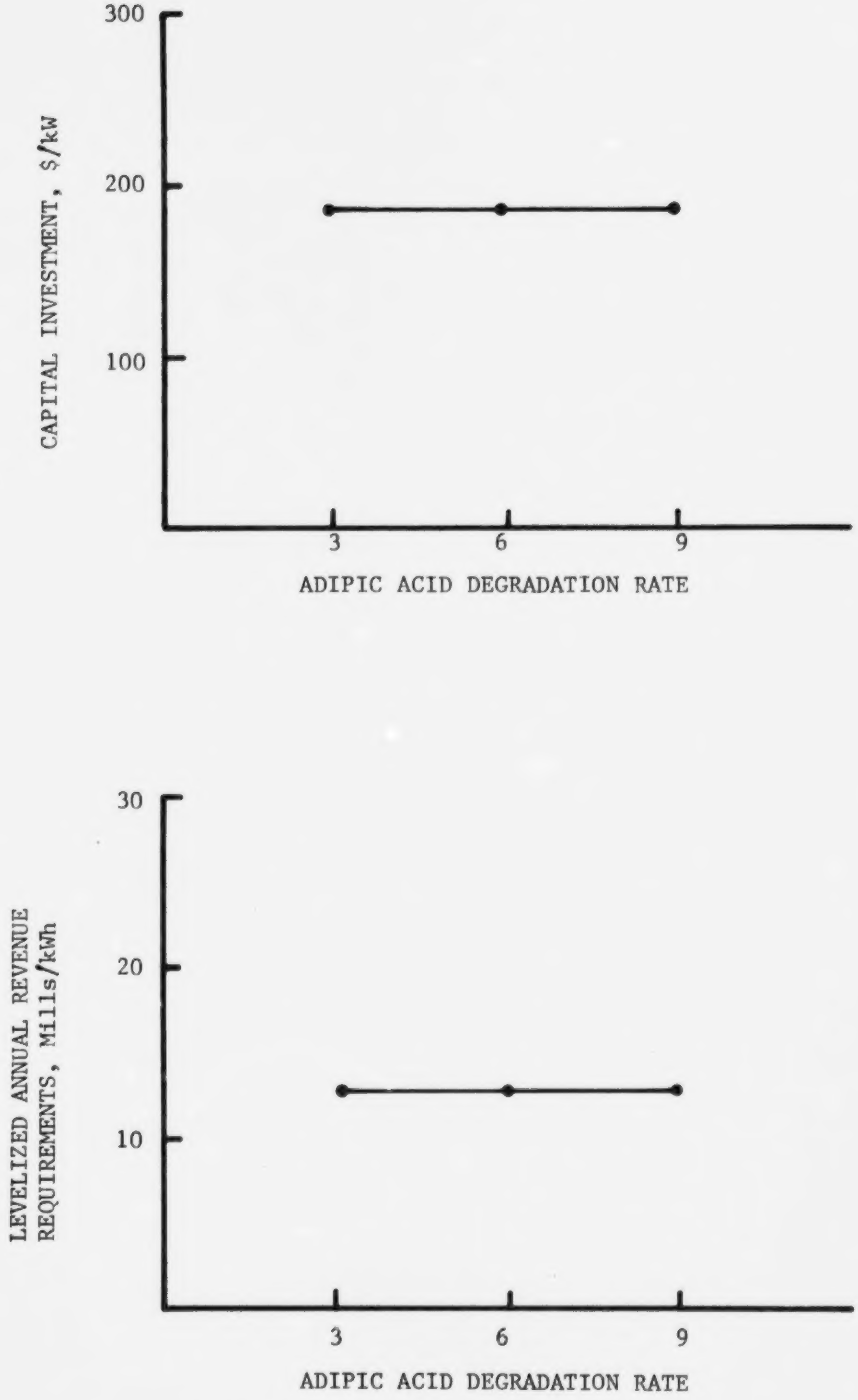


Figure S-9. Sensitivity of limestone adipic acid forced-oxidation process costs to the adipic acid degradation rate.

CONCLUSIONS

Processes with forced oxidation and landfill disposal have lower capital investments than processes without forced oxidation and pond disposal because the aggregate costs of forced oxidation and dewatering equipment and landfill construction are small, compared with the cost of pond construction. First-year annual revenue requirements are slightly higher for forced-oxidation processes under most conditions evaluated. At large waste volumes, however, the first-year annual revenue requirements for processes without forced oxidation are slightly lower. Processes without forced oxidation have slightly lower levelized annual revenue requirements in all cases.

The use of additives reduces both the capital investment and annual revenue requirements of limestone processes. For processes with forced oxidation and landfill disposal, the effects of magnesium oxide are marginal but with adipic acid more substantial reductions of 10% to 13% in both capital investment and annual revenue requirements are obtained. The cost reductions are reflected in capital investment by reduced equipment sizes and smaller waste volumes and in annual revenue requirements by reduced capital charges (the major factor) and other costs related to capital investment and by reduced absorbent and electricity costs. The larger cost reductions obtained with adipic acid are largely a result of a substantial improvement in limestone utilization. The costs related to the use of additives are a very small part of the total costs and both capital investment and annual revenue requirements are very insensitive to changes in the cost and consumption rate of the additives or to the use of lower cost substitutes. The higher removal efficiencies attainable with adipic acid also permit the use of the more economical partial scrubbing with bypass over a wider range of coal sulfur contents. For processes that use pond disposal, the use of additives has the same effect on capital investment as it does for processes using landfill disposal but the cost reduction in annual revenue requirements is reduced about one-half because of the additional additive loss in the low-solids waste.

Processes using TCA's are about 5% less expensive than processes using spray towers and 12% less expensive than processes using venturi - spray towers because of the smaller size and lower L/G ratio of the high-pressure-drop TCA, assuming that maintenance costs are the same percentage of equipment costs for all the processes. Changes of 15% in maintenance costs for either process make the annual revenue requirements of processes using TCA's and spray towers equivalent.

Economically, for unenhanced limestone scrubbing systems, the optimum absorber design favors higher limestone stoichiometries and lower L/G ratios. For the same SO_2 removal efficiency, a 1.4 stoichiometry and a low L/G is 14% to 18% lower in capital investment and annual revenue requirements than a design with a 1.1 stoichiometry and a high L/G ratio.

Partial scrubbing at 90% removal efficiency with bypass of some of the flue gas is more economical at coal sulfur contents below about 2% because the bypassed flue gas reduces reheat costs.

Lime processes have lower capital investments than limestone processes in all of the cases evaluated. They also have lower annual revenue requirements at lower coal sulfur contents. Lime processes are more sensitive to absorbent costs, however, and also to levelizing, since operating and maintenance costs constitute a larger percentage of the total annual revenue requirements. As a result, lime process annual revenue requirements increase more rapidly with coal sulfur content than those of limestone processes, particularly if levelized annual revenue requirements are compared. At a coal sulfur content of about 4.5% the first-year annual revenue requirements of lime and limestone processes are equal; at a coal sulfur content of about 3.0% the levelized annual revenue requirements are equal. As the coal sulfur content further increases limestone processes become increasingly more economical.

The largest components of annual revenue requirements for limestone processes, regardless of type, are, in order of decreasing importance, capital charges, maintenance, overheads, electricity, steam, limestone, and labor. The annual revenue requirements are quite sensitive to changes in capital investment, inflation, and the discount rate; moderately sensitive to maintenance costs; and relatively insensitive to changes in the other cost elements. The largest components of annual revenue requirements for lime processes are, in order of decreasing importance, capital charges, absorbent, maintenance, overheads, electricity, steam, and labor. The annual revenue requirements of lime processes are also most sensitive to changes in capital investment, inflation, and the discount rate. It is, however, moderately sensitive to absorbent cost as well as maintenance costs.

The costs of the base-case limestone process used in this study are approximately double those of the base-case limestone process used in similar previous studies, as a result of updated design and economic premises used in this study. The largest cost increase, about 25% of the total, is the result of advancing the cost index year from 1979 to 1982 for capital investment and 1980 to 1984 for annual revenue requirements. Incorporation of spray towers, spare absorbers, and emergency bypass produced the second largest increase, also about 25% of the total. General revisions of the economic premises account for about 15% of the increase. The remaining premise changes (incorporation of forced oxidation, substitution of landfill for pond disposal, and substitution of 1979 NSPS for 1971 NSPS) have a minor effect and, individually, do not account for more than a 5% increase in costs.

ECONOMIC EVALUATION OF ADVANCED LIMESTONE AND LIME FLUE GAS DESULFURIZATION PROCESSES

INTRODUCTION

This study investigates the economics of advanced limestone and lime wet scrubbing flue gas desulfurization (FGD) processes that incorporate recent (1975 to 1980) technological developments. These developments include forced oxidation, refined absorber design and operating conditions, landfill waste disposal, and the use of additives to enhance scrubbing efficiency. Most of the processes are based on EPA-performed or EPA-sponsored FGD evaluations, particularly those at the EPA Alkali Scrubbing Shawnee Test Facility at the Shawnee Steam Plant near Paducah, Kentucky. The economic results were obtained using a computer model developed by TVA and Bechtel National, Inc., from the Shawnee test facility data. In addition, this study is the first comprehensive limestone and lime FGD economic study by TVA to incorporate EPA-TVA design and economic premises that were revised in 1980 to reflect regulatory, technological, and economic conditions projected for the 1980's.

TVA has conducted economic evaluations of utility FGD processes for EPA since 1967. A large portion of these activities has consisted of comparative economic evaluations of conceptual designs based on current or evolving FGD technology. For these evaluations, design and economic premises were developed based on current and projected utility design and operating conditions and economic practices, and on projected economic conditions. The premises are designed to provide an equitable basis of comparison representative of practices and conditions anticipated for utility FGD in the near future.

Many of these studies have involved limestone and lime wet scrubbing FGD, both to keep abreast of evolving technology in this field and to provide a standard with which the economics of other processes could be compared. During this time several versions of limestone and lime FGD processes were evaluated, differing in design and operating conditions to represent current technology. In addition, power plant, fuel, and economic premises required periodic revisions.

Although data from many sources are used to develop conceptual designs for the limestone and lime FGD processes evaluated, results of tests at the Shawnee test facility have played an increasingly important part. The test facility was operated by TVA with Bechtel National, Inc., serving as technical director for EPA. Funding for operation of the test facility during limestone and lime wet scrubbing FGD testing from the late 1960's until 1980 was provided by EPA. Results of the tests were used to develop design and operating conditions for the limestone and lime processes using the EPA-TVA conceptual design economic premises.

In addition, as an adjunct of the Shawnee test facility program, Bechtel and TVA developed a computer model for projection of limestone and lime FGD costs based on the Shawnee test facility results. The model projects design and costs based on inputs such as power unit and fuel specifications, SO_2 removal efficiency requirements, absorber design specifications, and other variables depending on the requirements of the analyses. The design and economic criteria in the model are identical to the EPA-TVA design and economic premises used in the other conceptual design and economic studies.

Previous studies in the current series of TVA FGD economic evaluations for EPA, including two with limestone and lime processes (1,2), used premises developed during the mid-1970's. The premises were updated in 1980 to reflect current regulatory conditions, power unit design and operating procedures, FGD technology, and economic conditions projected into the early- to mid-1980's. The revisions resulted in an appreciable change in the projected costs of FGD processes compared with those of the previously used premises. These differences are also analyzed by comparison of base-case limestone FGD process costs developed using both premises.

BACKGROUND

The current limestone and lime wet scrubbing FGD processes such as those evaluated in this study largely evolved during the 1970's from processes which sometimes differed in fundamental concepts from the current ones. Much of the work in the early 1960's was conducted as private industrial studies and is sparsely documented in available literature. Earlier antecedents of FGD in the United States and Great Britain are well documented and provide perspective although their bearing on the development of current FGD technology may be arguably indirect.

The growth of limestone and lime FGD technology leading to the current processes is characterized by the increasing efficiency, reliability, and versatility attributable to advances of many types. Of particular importance from the perspective of this study is the increased quantification of the chemistry, which bears directly on the economics as it affects design features, and as it has led to modifications such as use of additives and forced oxidation. The chemistry has also been important, again bearing directly on economics, in an increased understanding of severe operating problems such as scaling. A growing understanding of the behavior of construction materials in FGD applications has also had important economic implications.

Disposal of wastes has an important effect on the economics of limestone and lime FGD. Developments in sludge dewatering, stabilization, and disposal techniques decisively affect a major cost area of these processes. Forced oxidation to gypsum, a major focus of this study, is an important part of these developments.

EARLY LIMESTONE AND LIME FGD DEVELOPMENTS

The antecedents of the current genre of limestone and lime FGD scrubbing processes have been reviewed to various depth in numerous publications. Discussions of work before the 1950's consist largely of applications other than steam-electric plants, foreign applications, and sporadic studies in the United States which did not lead directly to limestone and lime scrubbing (3). Direct scrubbing with slurries of limestone or lime is in some senses a relative latecomer to the technology of FGD. Two essentially unrelated areas of early studies are frequently cited in discussions of the precursors of current FGD practice. The work of Johnstone and his colleagues, which spans a period from the 1920's to the 1950's, is usually cited as the paramount example of early FGD studies in the United States. The use of limestone and lime wet scrubbing in commercial utility applications in England beginning in the 1930's is also widely mentioned in current FGD literature. The relationships of these early works to the present generation of limestone and lime FGD may be more apparent than genetic, however.

Johnstone did not, in fact, investigate limestone scrubbing although lime scrubbing was studied in some instances (4). Much of his work was directed toward technologies that would be economically acceptable under the prevailing conditions, in which environmental restraints played a smaller role than they do today. Johnstone regarded a slurry as too expensive for use with high-sulfur fuels (5). He directed much of his efforts toward improvements in water scrubbing (6), equipment evaluation (7), and regenerable processes (8). Johnstone's early work with catalysts, along with Copson's (9), is credited as a basis for certain aspects of later double-alkali processes (10).

The early development of utility FGD in England grew from the necessity of coping with atmospheric conditions created by large single-site sources, which by the 1920's had become a subject of public concern. That portion of this concern directed toward the utility industry in the form of legal and licensing restrictions (3) led to the use of full-scale scrubbing systems using river water, sometimes augmented with limestone, and to the development of a lime and limestone (chalk) process. The latter was developed on a pilot scale by James Howden & Company and Imperial Chemical Industries (I.C.I.) and later used at two power plants (11). Specific aspects of this phase in the development of FGD are frequently cited in current literature. It has also been summarized in several reviews of FGD technology (12). Most of the original literature occurs in British trade journals and commission reports which are not readily available in the United States and which have apparently not been systematically reported. The use of the Howden-I.C.I. process did not long survive World War II and other scrubbing installations were gradually abandoned. The use of utility FGD is no longer considered an applicable technology in Great Britain (13).

Much of the British experience with FGD appears a microcosm of later U.S. experience. Most of the same problems with plugging, scaling, and corrosion (14), many of the same process procedures, and some chemical results are common to both (15). To what extent the British experience influenced subsequent U.S. development of limestone and lime FGD has not been elucidated. Most comparisons of common phenomena in the literature are historical.

Another, more direct, lineage of current limestone and lime scrubbing can be found in the extensive efforts of the 1960's to develop dry injection FGD. One of the most extensively described dry injection studies in open literature was the EPA-sponsored testing at TVA's Shawnee Steam Plant (16). The economic attractiveness of boiler injection led to other full-scale installations (17), all of which failed because of fouling in the boiler and low removal efficiencies. Use of wet scrubbers downstream of the boiler to collect the reaction byproducts and provide more contact between the flue gas and the limestone was incorporated into these processes to improve removal efficiencies. Development of wet scrubbing processes with slurry addition in the scrubber system paralleled these studies but on a smaller scale. Although dry injection ("front end" processes) remained in active consideration in the early 1970's, slurry addition in the scrubber system ("tail end" processes) rapidly gained dominance (18).

Throughout this period of its evolution limestone and lime scrubbing were somewhat lightly regarded in relation to more sophisticated processes. In addition to Johnstone's disinterest, summaries of progress in utility FGD mention limestone and lime scrubbing only in passing (19,20). Even the burgeoning development of the early 1970's was regarded in some respects as a practical measure permitting the development of more desirable processes.

The limestone and lime FGD processes evolved from these studies are, in their main elements, the processes in use today. There remained, however, an extensive need for refinement of design and operating concepts to solve problems general, and sometimes fatal, to early designs. Among these were corrosion and scaling and plugging, the solution of which occupied much of the early development work. In addition, extensive development was needed in improvement of removal efficiency, absorbent utilization, water use, and waste management.

In many respects an expanded knowledge of the process chemistry was central to the resolution of many difficulties and deficiencies of limestone and lime scrubbing. Through it scaling and plugging control was improved, removal efficiency increased, and absorbent utilization increased. This allowed (along with a concomitant increase in performance, process control, and corrosion data) improved designs leading to more efficient and reliable processes, and better control of waste properties.

Although the present day limestone and lime processes are more complex in comparison with early processes, they have also become simpler in some respects such as absorber design. The advantage of the more reactive lime over limestone has also been reduced and the possible methods of waste disposal expanded.

CHEMISTRY OF ABSORPTION

The chemistry of limestone and lime scrubbing FGD is widely described in FGD literature, although frequently in simplified styles that omit or obscure important aspects. More extensive treatments are provided by Lowell (21), the reports of the Shawnee test facility advanced program (22-27), Rochelle (28), Rochelle and King (29), and Corbett (30).

Simplified SO_2-CO_2 Systems

It is useful to conceptualize the overall chemistry within the structure of the SO_2-CO_2-H_2O gas-liquid equilibrium relationships, which can be used at least qualitatively to define the reactions occurring in the scrubbing loop. In particular, recognition of the roles of pH and common ion equilibrium as definitive properties is important in visualizing the relationships of other components in the system.

Sulfur dioxide is a readily soluble gas, compared with most gases present in flue gas. At one atmosphere and 25°C, for example, the equilibrium concentration of SO_2 in water is 1.5M, compared with 0.03M for CO_2 and 0.001M for O_2 (31). The partial pressures of these gases in a typical flue gas differ greatly, however. For the base-case conditions in this study, the molar concentrations of SO_2, CO_2, and O_2 are 0.2%, 12.1%, and 5.5% respectively. Using the respective Henry's law constants, the driving force for CO_2 absorption is ideally somewhat greater than for SO_2 absorption.

Although represented in many forms, the equilibria defining the gas-liquid relationships of the solution commonly called sulfurous acid can be written in simplified form as:

$$SO_2(g) \rightleftarrows SO_2 \cdot H_2O(aq) \qquad H \approx 0.03$$

$$SO_2 \cdot H_2O + H_2O \rightleftarrows HSO_3^- + H_3O^+ \qquad K_1 \approx 10^{-2}$$

$$HSO_3^- + H_2O \rightleftarrows SO_3^{2-} + H_3O^+ \qquad K_2 \approx 10^{-7}$$

$$H_2O + H_2O \rightleftarrows H_3O^+ + OH^- \qquad K \approx 10^{-14}$$

Where H is the Henry's law constant, P_{gas}/P_{liquid}, and K is the dissociation constant.

Although the state of undissociated SO_2 in solution is often represented as H_2SO_3, this species has not been identified in solution (32). The actual state is generally accepted as a loosely hydrated molecule (33). The bisulfite form exists only in dilute solutions (34). Other species such as the disulfite also occur at very low concentrations but are usually ignored in calculations.

The absorption and dissociation of CO_2 in water can be expressed in an analogous series of equations by substituting carbon for sulfur and using the appropriate Henry's law and equilibrium constants. In this case H_2CO_3 is a real species although most of the unhydrolyzed CO_2 also exists as an aqua-complex (35). The approximate Henry's law and equilibrium constants are $H \approx 1.2$, $K_1 \approx 10^{-6}$, and $K_2 \approx 10^{-10}$.

The liquid composition in terms of the relative concentrations of the dissolved species at different pH's can be calculated for SO_2 and CO_2 using the dissociation constants. Plots of these calculations are sometimes shown in the literature to illustrate the importance of the various species in limestone and lime scrubbing pH ranges, such as the diagrams shown in Figure 1. As shown, $SO_3 \cdot H_2O$ and CO_3^{2-} play a minor role in equilibrium relationships in the pH range of limestone and lime scrubbing. Also, since both the SO_2 and CO_2 systems are buffers, the pH range has a direct

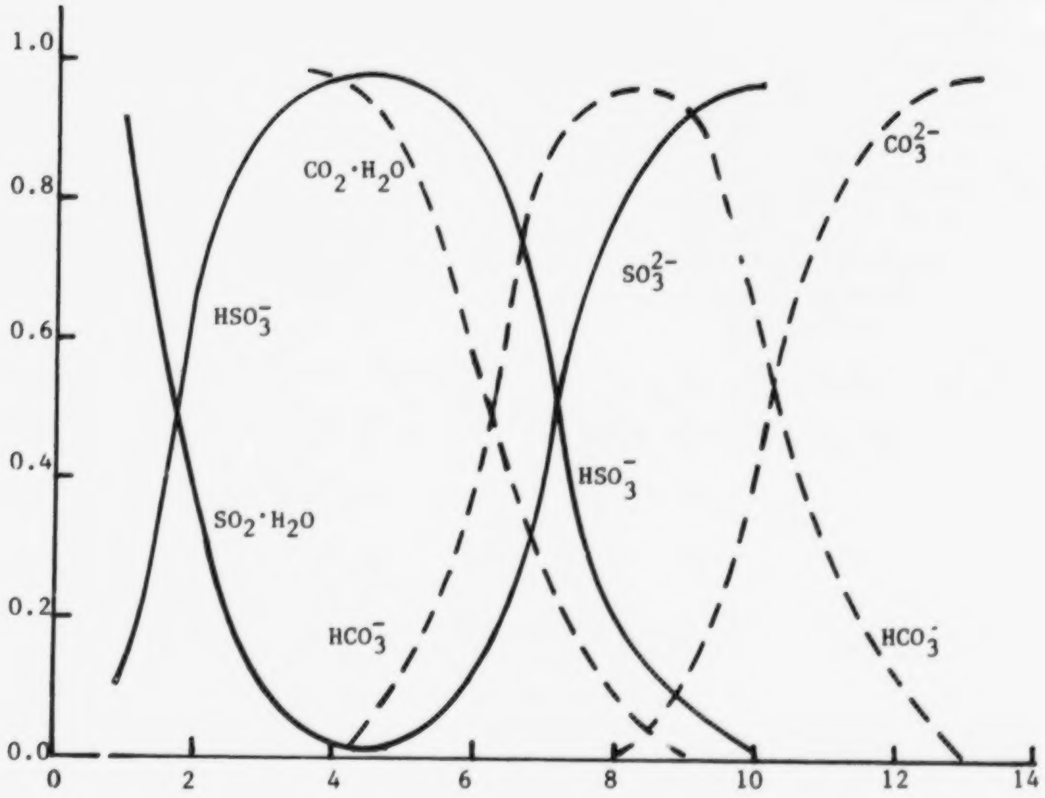


Figure 1. Relationship of sulfur dioxide and carbonate species in solution as a function of pH.

bearing on the absorption efficiency and on the value of pH as a process control criterion.

The role of flue gas CO_2 in producing basic dissolved species is not in itself either sufficient or important in removal of SO_2 from the flue gas. The equilibrium reactions do, however, serve as the basis of the chemical description in which limestone or lime is used as a means of increasing the quantity of basic species in solution and as a source of calcium ions to remove sulfur ions from solution.

In addition to these basic reactions, oxidation of sulfite and the presence of other dissolved species also play important roles in the overall scrubbing process. The normal and unavoidable oxidation of some of the sulfite in solution by O_2 absorbed from the flue gas, and possibly from other sources in the system, can range from a few percent to well over 50% of the SO_2 absorbed for flue gases from low-sulfur coals. Since the O_2 absorption rate is not directly dependent on the SO_2 absorption rate, low SO_2 concentrations in the flue gas will result in higher ratios of O_2 to SO_2 absorbed and higher oxidation rates. The O_2 absorption rate can be controlled to some extent by controlling gas-liquid contact within the bounds dictated by SO_2 removal requirements and by increasing the slurry ionic concentration. In some double-alkali and byproduct-producing FGD systems sulfate formation is a problem resulting in special processing steps and loss of raw materials (36).

As a somewhat weaker acid than sulfurous acid, carbonic acid acts as a base in the Lewis acid base sense (37). The carbonate species are proton acceptors for the sulfite species, as is SO_3^{2-} for the hydrated SO_2 molecule:

$$SO_2 \cdot H_2O + HCO_3^- \rightleftarrows HSO_3^- + CO_2 \cdot H_2O$$

$$HSO_3^- + HCO_3^- \rightleftarrows SO_3^{2-} + CO_2 \cdot H_2O$$

$$SO_2 \cdot H_2O + SO_3^{2-} \rightleftarrows 2HSO_3^-$$

It is common to regard the various species in solution that act as Lewis bases to sulfurous acid as total dissolved alkalinity, thus identifying the total equivalents of base that will act as proton acceptors for the SO_2 absorbed. These include the carbonate species, OH^-, and SO_3^{2-} as well as the various ion pairs such as $CaSO_3^0$ and radicals such as $CaOH^{+}\cdot$. The reactions among these species are normally regarded as instantaneous in terms of gross system phenomena although rates may have a critical role in liquid-film-controlled absorption processes. The oxidation of sulfite is also largely a rapid scrubber regimen reaction (30).

Dissolution and Precipitation Reactions

In contrast, some dissolution and crystallization rates are slow in terms of both absorber and liquid loop residence times. Limestone dissolution rates vary widely depending on the composition, texture, and particle size. In terms of approach to equilibrium, quantified by pH changes in scrubbing liquids (38), the dissolution rate is measured in terms of hours. Similarly, a considerable degree of calcium sulfite and sulfate supersaturation is a common and important phenomenon of scrubbing liquids (30).

Calcite, the reactive component of most limestones, is sparingly soluble in water, as shown in the simplified pH-dependent equilibrium:

$$CaCO_3 + H_2O \rightleftarrows Ca^{2+} + HCO_3^- + OH^- \qquad K \approx 8$$

Its solubility is ultimately controlled by the partial pressure of CO_2 in the flue gas.

Limestone dissolution rates and the factors affecting them have been the subject of numerous investigations, both as independent studies (39-40) and as a part of the evaluations at the Shawnee test facility (26). It is affected by various physical and chemical properties such as porosity, crystal habit, and the degree of dolomitic composition as well as the particle size as determined by the fineness of grind. Normally the desired dissolution rates require a stoichiometric excess of $CaCO_3$ of 1.1 to 1.5 (all stoichiometries are expressed as mols of $CaCO_3$ added per mol of SO_2 absorbed) though the ratio is highly dependent on the combined effects of removal efficiency, design, and operating conditions. These numerous interactions have led to differing conclusions on the effects of stoichiometric ratio (30).

Because of the slow dissolution rate, limestone dissolution occurs in the absorber as well as the hold tank in most limestone FGD systems. In some cases absorber dissolution accounts for a major portion of the total limestone reacted (27). This is unique to simple limestone scrubbing and leads to particular characteristics of the process.

The degree to which limestone dissociates in the absorber for a particular system is a function of the limestone stoichiometry. At high stoichiometries (over about 1.3) the pH of the entering slurry is high and there is abundant solid $CaCO_3$ available. Under such conditions limestone dissolution may provide the major portion of the dissolved alkalinity. At lower stoichiometries limestone dissolution becomes less important and $CaSO_3$ dissolution more important as a source of dissolved alkalinity. The importance of limestone dissolution in the absorber lies in the increase in Ca^{2+} concentration in the absorber. If limestone dissolution is the major source, the ratio of Ca^{2+} to SO_2 absorbed can approach 0.5, according to the reaction:

$$CaCO_3 + 2SO_2 + H_2O \rightarrow Ca^{2+} + 2HSO_3^- + CO_2$$

Conversely, if $CaSO_3$ is the major source, the ratio can approach 1.0:

$$CaSO_3 + SO_2 + H_2O \rightarrow Ca^{2+} + 2HSO_3^-$$

Dissociation of $CaCO_3$ is thus advantageous from the standpoint of $CaSO_4$ saturation because the increase in Ca^{2+} concentration in the absorber is less but the stoichiometries required may be undesirable for operational and economic reasons. Conversely, dominate $CaSO_3$ dissociation would require a high degree of $CaCO_3$ dissociation in the external liquid loop with impractical hold tank residence times. Either extreme might also involve pH changes in the absorber that would be undesirable.

A rapid decrease in pH, of course, increases the limestone dissolution rate. If the pH decreases too rapidly, however, limestone blinding may occur. This phenomenon, discussed as an operating problem in the Shawnee test facility reports (27), has been described as a liquid-film phenomenon related to the limestone dissolution rate (28). At high dissolution rates (low bulk liquid pH), the liquid surrounding the particle reaches a high concentration of Ca^{2+} in which $CaSO_3$ is precipitated rapidly enough to nucleate on and isolate the particles.

In addition to its effect on limestone dissolution, the extent to which the pH decreases in the absorber also has important effects on the formation of solid $CaSO_3$ since at lower pH's a larger proportion of the sulfur species is in the bisulfite form. If the pH change is small, $CaSO_3$ may precipitate in the absorber, possibly creating plugging problems.

In contrast to $CaSO_3$ solubility, $CaSO_4$ solubility is relatively independent of pH. Precipitation of $CaSO_4$ in the absorber is essentially dependent on the inlet concentration, the oxidation rate, the presence of seed crystals, and the increase in Ca^{2+} concentration in the absorber. The phenomenon has been extensively investigated because of its relation to scale formation (41). Because of its propensity to form supersaturated solutions, $CaSO_4$ can increase to highly supersaturated levels at which relatively sudden and intensive crystal nucleation occurs (30). The degree of supersaturation is reduced at oxidation levels below about 20% by removal of the sulfate as a component of $CaSO_3 \cdot 1/2H_2O$ crystals, a mechanism long observed but only identified in the past few years (15). It is sometimes described as a solid solution phenomenon (30).

In limestone and lime systems the principal problem of $CaSO_4$ supersaturation is sulfate scale formation, one of the major operating problems of limestone and lime FGD, which is discussed in more detail below. From the aspect of SO_2 scrubbing chemistry $CaSO_4$, being much more soluble than the corresponding $CaSO_3$ species, increases the calcium ion concentration. This induces a corresponding decrease in basic sulfite species, to meet solubility equilibrium requirements. The sulfate thus acts as a control of dissolved alkalinity.

Overall Chemistry

In this most simplified version of basic limestone scrubbing chemistry there are thus three gaseous components, eight possible aqueous components, and three solid components which occupy two primary regimens, the absorber and the external liquid loop. Important chemical factors within the absorber are primarily related to solid-liquid considerations as controlled by the amount of solid $CaCO_3$, $CaSO_3$, and $CaSO_4$ present, the pH, and the quantity of SO_2 absorbed in relation to the quantity of the liquid.

Control of solids formation in the absorber is a major consideration in limestone and lime scrubbing. Control of $CaCO_3$ and $CaSO_3$ formation is largely a matter of controlling the maximum inlet and outlet pH. This decreases the tendency of the incoming liquid to precipitate $CaCO_3$ and for a small pH change to precipitate $CaSO_3$ as SO_2 is absorbed. At the Shawnee test facility maximum inlet pH's of about 6.0 and 8.0 for limestone and lime scrubbing, respectively, and outlet pH's of about 5.5 for both were found satisfactory (27).

Other Ionic Species

The presence of other ionic species in solution is also of paramount importance in the overall relationships of the basic SO_2-CO_2-$CaCO_3$-H_2O system equilibrium relationships. Some ion pairs and complexes in the simple system itself such as $CaSO_3{}^0$, $CaCO_3{}^0$, and $CaOH^+$ which are basic with respect to $SO_2 \cdot H_2O$ affect the chemical relationships. Many other substances, which accumulate in the scrubbing liquid as components of the flue gas or raw materials or as substances deliberately added, are important. Primary among those most important in limestone and lime scrubbing are Cl^- from the flue gas and Mg^{2+} either as a limestone component or as a separate additive. Similarly, buffers such as adipic acid can also be used to alter the simple equilibrium relationships.

Chloride, which enters the scrubbing liquid from the flue gas as highly soluble HCl, is sometimes the major dissolved component in scrubbing liquids, ranging upward to 10,000 to 20,000 ppm, particularly in processes recycling a large portion of the process water (42). The chemical effect of chloride is to increase the concentration of calcium ion in solution since equilibria involving the calcium ion must also satisfy the equilibrium of highly soluble $CaCl_2$. The overall result is a decrease in dissolved basic species such as $SO_3{}^{2-}$ and a concomitant decrease in pH, both of which decrease scrubbing efficiency.

Lime Processes

The use of lime produces little change in the basic chemistry of the process but considerable change in the overall kinetics. The introduction of the more rapidly dissociating hydroxide results in a higher pH and essentially complete dissociation of $Ca(OH)_2$ in the hold tank instead of the absorber. The higher pH's attainable increase the concentration of alkaline species in the liquid, thus increasing the scrubbing efficiency. At equal pH's, however, dissolved alkalinities in both lime and limestone are essentially equal (25,30). Carbonate species, formed from absorbed CO_2, also have a role in lime scrubbing but the extent to which $CaCO_3$ dissolution in the absorber contributes to SO_2 removal is not well defined. The dissolution of $CaSO_3$ is believed to be the major source of basic anions in the absorber. Precipitated $CaCO_3$ lost in the waste stream is the major alkalinity loss that limits the absorbent utilization efficiency, which because of the rapid dissolution of $Ca(OH)_2$ is higher (1.0 to 1.2 stoichiometry) than limestone utilization.

OXIDATION OF SULFITE

The oxidation of the sulfite ion in solution to sulfate is one of the classic chain reactions of chemistry. In addition, it is of practical importance in many fields of technology such as atmospheric chemistry, food processing, numerous industrial processes, and in the empirical development of mass transfer theories. This broad interest has generated, for a period of over 100 years, a voluminous literature on the reaction and the many factors affecting it. In spite of this, a general uncertainty still surrounds many facets of the reaction and to a lesser extent the chain sequence itself.

The oxidation mechanism is a typical chain reaction involving various transient species of sulfur ions. The main features of the reaction first proposed by Backstrom in the 1920's and 1930's remain the basis of most representations, such as that after Hegg and Hobbs (43).

$$SO_3^{2-} + M^+ \rightarrow SO_3^-\cdot + M\cdot$$

$$SO_3^-\cdot + O_2 \rightarrow SO_5^-\cdot$$

$$SO_5^-\cdot + HSO_3^- \rightarrow HSO_5^- + SO_3^-\cdot \qquad pH < 7$$

$$SO_5^-\cdot + SO_3^{2-} \rightarrow SO_5^{2-} + SO_3^-\cdot \qquad pH > 7$$

$$SO_3^{2-} + HSO_3^- \rightarrow HSO_4 + SO_4^{2-} \qquad pH < 7$$

$$SO_3^{2-} + SO_5^{2-} \rightarrow 2SO_4^{2-} \qquad pH > 7$$

$$SO_3^-\cdot + SO_5^-\cdot \rightarrow S_2O_6^{2-} + O_2$$

$$SO_5^-\cdot + \text{inhibitor} \rightarrow \text{inert products}$$

In the sequence shown, the initiator is a metal ion such as hexaquo cobalt that abstracts an electron from the SO_3^{2-} ion to produce the active center $SO_3^{-\cdot}$. Other investigators (44,45) provide similar reactions in which the mechanism proceeds through a free radical chain involving $SO_3^{-\cdot}$, SO_5^{2-}, and $SO_5^{-\cdot}$. Hayon and others (46) propose a similar mechanism with the inclusion of an $SO_4^{-\cdot}$ radical that they infer, although it was not observed, because of the inhibitive effect of alcohols. These authors also discuss numerous earlier studies. There is still lively discussion on many aspects of the reaction such as the existence and role of particular intermediates, the role of HSO_3^- and SO_3^{2-} as oxygen carriers (pH dependence), the order of the reaction with respect to the participating species, and the effects of the physical conditions under which the reaction occurs.

The reaction is readily induced by photolysis or by a number of metallic and nonmetallic catalysts. There are also many inhibitors to the reaction. In some cases the same material may serve as a catalyst or an inhibitor depending on the physical or chemical conditions of the solution. It is the catalyst-induced reaction that has received the most extensive study and which is most applicable to FGD.

A voluminous literature exists on deliberately catalyzed and uncatalyzed aqueous sulfite oxidation in the field of atmospheric chemistry (47,48,49). These studies are interesting because of their similarity to FGD oxidation in some respects: they deal with dilute solutions of eclectic composition which is not subject to the controls of laboratory studies.

Absorption of oxygen from gases into sulfite solutions containing a catalyst has been used for many years as a method for determining interfacial areas and determining film characteristics in gas-liquid absorption devices. The focus of interest in these studies is often on methodology and kinetics. Also, they are frequently carried out at high oxygen and sulfite concentrations and high pH's, relative to normal FGD conditions. This area of study has, however, produced a considerable body of data on reaction rates, catalysts, and inhibitors directly or generally applicable to FGD.

There has been rather poor agreement in these studies on reaction orders and rate constants, often attributed to the purity of the solutions (50) and the chemicals used (51). It is generally recognized that the roles of pH, temperature, and concentrations are important (52) and that rate studies unaffected by catalysts and inhibitors are difficult to achieve (53).

Catalysts and Inhibitors

No topic of the sulfite oxidation mechanism has, in fact, received as much attention as the role of catalysts and inhibitors, both of which may have large effects on the rate. In spite of extensive studies, which extend back to the 19th century and which have generated an immense literature, these still remain important regions of question in many areas of application. It has been long recognized that numerous metals and some nonmetallic compounds act as catalysts and others as inhibitors of the oxidation reaction. The work of Johnstone (54) and Alyea and Backstrom (55) is typical of early investigations. Hudson (56) and Hudson and others (53) provide extensive compilations of studies directly or indirectly applicable to FGD.

Most common metallic elements have been evaluated as catalysts for aqueous sulfite oxidation but the primary attention has focused on the elements of the first transition series (which includes copper in its normal oxidation states). Manganese, iron, cobalt, and copper have received the most attention and have proven the most effective in most cases. Nitrogen dioxide has also been extensively studied as a sulfite oxidation catalyst or chain initiator (57,58), as have the hydroxyl and perchlorate ions. Much of this work has taken place in the fields of gas-liquid mass transfer and atmospheric chemistry and thus spans a wide range of concentrations and pH's not frequently comparable to FGD conditions.

Certain organic substances have long been recognized as inhibitors of aqueous sulfite oxidation. Alyea and Backstrom (55) studied the effects of alcohols, attributing the inhibition to a reaction chain interruption in which the alcohol is preferentially oxidized. Phenols are most frequently cited as examples of inhibitors; less frequently other classes of hydroxyl containing organic compounds, including rubber, are cited. Hudson (56) calls rubber "the bane of much experimental work . . . in apparatus . . . ." Hudson experimented briefly with polycarboxylic acids, including adipic, in which he found an inhibiting effect in all cases, particularly with some acids containing alcoholic hydroxyl groups. He concluded that the alcoholic hydroxyl is the primary active group, though sometimes hindered by steric effects, and the carboxylic hydroxyl group is a less effective active group.

Manvelyan and others (59), on the other hand, found p-phenylenediamine to be a more powerful inhibitor in $MgSO_3$ solutions than hydroxyl-bearing inhibitors. Other groups such as amines, amides, aldehydes, and ketones have, in fact, been cited as possible inhibitors (43). de Wall and Okeson (51), using cobalt as a catalyst, found phenol to have no effect. Johnstone found copper to be a powerful inhibitor in his studies (54) although copper has been used as a catalyst in other studies under different conditions.

The effect of inhibitors thus appears diverse and extensive. The effect of organic compounds bearing carboxylic hydroxyls appears well established as does similar noncarboxylic oxygen-bearing compounds. To what extent these are effective under different chemical and physical conditions is not well established, however. The effect of inorganic oxygen-bearing compounds such as NO may be similar, a preferential oxidation mechanism that interrupts the chain sequence. The role of metals such as copper is well established but, again, not well defined in terms of chemical and physical conditions. In most cases, particularly in the case of organic materials, the inhibitor is probably consumed in the reaction, further complicating assessment of the process. It is imprecise to refer to such inhibitors as catalyst poisons. Chelating agents have, however, a demonstrated effect in reducing the effectiveness of metallic catalysts (56).

In FGD most early catalyst studies were directed at increasing the absorption of SO_2 by water by reducing the sulfite ion concentration of the scrubbing liquid. Work by Johnstone (54), Copson (9), and Tarbutton (60) are typical and provide access to other studies. Manganese was the primary catalyst and the evaluations were generally successful in increasing absorption efficiency though not to the extent that a commercially useful acid concentration (the primary objective) was attained. This work did, however, serve as a partial basis of later double-alkali gypsum processes (10). Also significant in these studies was the recognition of the importance of inhibitors.

Typical flue gas compositions provide abundant materials for numerous combinations of known catalysts and inhibitors in the scrubbing liquid. In addition, the use of limestone and lime absorbents and buffer additives introduces additional chemical variables. Not withstanding these complexities, considerable progress has been made in defining both the practical aspects of forced oxidation applied to FGD and the effects of chemical variables on a more theoretical basis.

Forced Oxidation FGD

Forced oxidation by sparging air into a bleedstream of the absorber loop, or into the liquid in the absorber loop itself, has been a fixture of many double-alkali FGD processes since the 1960's (61,62). In most cases the increase in pH, and the corresponding decrease in the solubility of sulfite species, accompanying bleedstream forced oxidation has required acidification of the slurry to attain high oxidation efficiencies (63). Consequently, forced oxidation in limestone and lime scrubbing processes was first accomplished in the scrubbing loop using a two-stage system, each stage having an independent circulating system. The second stage (relative to the flow of the flue gas) operates at higher pH's for high removal efficiency. The slurry from the second stage is then used in the first stage, which serves as the oxidation loop. Removal efficiency is poor in the first stage but sufficient SO_2 is absorbed to maintain a low pH to facilitate oxidation. The two-stage system permits pH control to maximize absorbent utilization and oxidation in the first stage and removal efficiency in the second stage. The two-stage, forced-oxidation evaluation by EPA at the Industrial Environmental Research Laboratory at Research Triangle Park, North Carolina (IERL-RTP) pilot plant and the Shawnee test facility has been extensively described (26,27,38). A similar two-stage, forced-oxidation process was developed by Research-Cottrell (64). The Chiyoda 121 process employs a special absorber that attains satisfactory single-tank oxidation (65,66).

A single-tank, forced-oxidation Combustion Engineering system is also used by Northern States Power (67). Single-stage, forced oxidation in limestone processes was also successfully evaluated at the IERL-RTP pilot plant (68,69). A single absorber and hold tank was used and air was sparged into the hold tank. In this case the high concentration of $CaSO_3$ in the liquid leaving the scrubber maintained efficient oxidation conditions in the hold tank. It was also found that aeration increased the pH of the slurry, presumably by purging CO_2 and thus reducing HCO_3^-, which enchances limestone dissolution. Whatever the cause, the overall effect was an improved removal efficiency. Magnesium was ineffective in improving removal efficiency, however, possibly because of the low level of sulfite in the liquid. These results were further quantified at the Shawnee test facility (26,27). The use of two hold tanks in series, one for oxidation and one for absorbent addition, was found to improve oxidation and removal efficiency.

In contrast, bleedstream oxidation proved difficult at the Shawnee test facility because of the pH rise unless magnesium additive was used to increase the sulfate solubility (26,27). It did, however, permit effective use of magnesium additive in the scrubber loop.

In general, oxidation levels of over 95% were attained in both systems at ph's of 4.5 to 5.5 in the oxidation tank with air stoichiometries of 1.5 to about 2.8 atoms of O per mol of SO_2 absorbed. Depending on the operating conditions, limestone utilizations of 80% to 98% were attained.

Other facets of the Shawnee test facility forced-oxidation tests were evaluations of air injection systems consisting of nozzle eductors and pipe grids (26). Slurry depths up to 18 feet were evaluated. Other studies at the IERL-RTP pilot plant as well as studies in the literature (69,70) suggest that orifice size is unimportant but that the liquid depth is of considerable significance in promoting oxygen mass transfer efficiency.

A related EPA-sponsored study by Hudson and his colleagues bears on literature studies, bench-scale evaluations, and mathematical characterization of forced oxidation applicable to FGD (53,56). Extensive reviews of existing literature were made and general absorption models defined. Further work directed specifically to limestone and lime scrubbing consisted of bench-scale investigations of reaction rates, mass transfer rates, and the effects of catalysts and organic acids. Extensive development of mathematical models based on the Bechtel-Radian computer model was also part of these studies.

ADDITIVE-ENHANCED SCRUBBING LIQUIDS

The efficiency-limiting solubilities of the basic species in limestone and lime scrubbing liquids immediately draw attention to the possibility of chemical modification of the solution to supplement or increase the concentrations of these species. One such approach, evident from chemical equilibria considerations, is the introduction of cations forming more soluble salts of the basic anions than calcium does. The use of a magnesium compound to supply the cation is attractive for a number of practical and economic reasons, as discussed below. A second somewhat different means of increasing the basic anion concentration of the liquid is the use of a buffer to supplement the basic anion concentration and to augment the existing buffering capacity of the HSO_3^-/SO_3^{2-} and $CO_2 \cdot H_2O/HCO_3^-$ pairs.

Magnesium-Enhanced Scrubbing Liquids

Addition of cations forming more soluble salts of the basic anions than calcium does results in an increase in the basic anions to meet the new equilibria requirements and a concomitant decrease in the calcium ion concentration. Many such cations can be used, among them Na^+, NH_3^+, and Mg^{2+}

(29). Mg^{2+} is particularly attractive, however. It is already present to some degree in limestone and lime scrubbing liquids, it does not form highly soluble salts that would contribute to disposal problems, inexpensive magnesium salts are readily available, and it forms ion pairs that contribute substantially to the total alkalinity of the liquid. The effect of magnesium lies in the equilibrium relationships such as:

$$MgSO_4 \rightleftarrows Mg^{2+} + SO_4^{2-} \qquad K = 10^{-3}$$

which must be accounted for in the overall equilibrium.

Simple algebraic manipulations demonstrate the overall effect, which is to increase the concentrations of $SO_4{}^{2-}$ and the various basic species such as $SO_3{}^-$ and $HCO_3{}^-$ and decrease the concentration of Ca^{2+}. The liquid thus becomes more alkaline with respect to sulfurous acid. Specific relationships are often illustrated, such as the relationship of $SO_3{}^{2-}$ and $HCO_3{}^-$ to $SO_4{}^{2-}$, the role of pH, and the effect of the partial pressure of CO_2.

For example, the various equilibria provide the relationships:

$$[SO_3^{2-}]/[SO_4^{2-}] = K_1/K_2$$

$$[HCO_3^-]/[SO_4^{2-}] = PCO_2(K_3/K_2)$$

where K_1, K_2, and K_3 are the respective dissociation constants of $CaSO_3$, $CaSO_4$, and $CaCO_3$. In an actual scrubbing liquid the determination of relationships must include the effects of ionic strengths, ion activities, and chemical complex and ion pair formations that lead to extensive theoretical and empirical extensions of these illustrative examples. Substantive empirical verification of such calculations is usually necessary. Qualitative discussions directly pertinent to limestone and lime scrubbing are provided by Head (25), Burbank and Wang (27), Rochelle (28), and Rochelle and King (29), among others.

A similar but obverse effect occurs with the introduction of anions that form stronger acids than sulfurous acid and that form more soluble calcium salts than the sulfite. The overall effect is an increase in calcium ion concentration and a decrease in basic anions and pH. The species of practical importance in limestone and lime scrubbing is chloride, the effects of which have been discussed. The sulfate ion is also such an anion, the effect of which is limited by the solubility of its calcium and magnesium compounds, compared with $CaCl_2$.

The deliberate introduction of magnesium did not appear to be recognized in early FGD work. Magnesium in the form of dolomite was sometimes regarded as a detriment to limestone dissolution. It was also considered undesirable because of the increased solubility of the waste (71).

The use of magnesium as a controlled, absorption-enhancing additive in limestone scrubbing was briefly investigated by TVA in 1970 (72). Several studies using high-magnesium lime were also conducted in the early 1970's. These led to the use of special limes (73) in prototype and full-scale FGD operations (74,75). Patents have been obtained on various aspects of these studies (76,77).

The use of magnesium as a controlled additive was evaluated in limestone and lime scrubbing at the EPA IERL-RTP pilot plant in 1973 by Borgwardt (15). Additional studies were subsequently conducted at the Shawnee test facility (25,26). Borgwardt's investigations used magnesium to control sulfate saturation levels in investigations of sulfate removal in $CaSO_3$ crystals. The effects of Mg^{2+}, Cl^-, and sulfate activity were quantified in these studies. In the tests at the Shawnee test facility, removal efficiencies were increased up to 15% with magnesium addition. Burbank and Wang (27) also summarize these studies and provide further discussion of the chemical aspects based on Shawnee test facility results.

The relationship of the opposing effects of magnesium and chloride was found to be additive for practical purposes, leading to the use of the term effective magnesium. Effective magnesium is determined by subtracting the equivalents of chloride in solution from the equivalents of magnesium present ([Mg] - 2[Cl]). At the Shawnee test facility, using effective magnesium levels of about 9,000 ppm, limestone scrubbing efficiency was increased about 15%. Lime scrubbing efficiency, already above 90%, was not improved within the range of Mg^{2+} concentration tested. Forced oxidation was not affected but magnesium enhancement was not effective in oxidized slurries with little sulfite present. Slurry settling rates were reduced but overall dewatering properties were unaffected. Bleedstream oxidation efficiency was also enhanced by the presence of magnesium.

Buffer Additives

Another somewhat different means of increasing the scrubbing efficiency of calcium-based liquids is the use of a buffer to augment the existing buffering capacities of the HSO_3^-/SO_3^{2-} and $CO_2 \cdot H_2O/HCO_3^-$ pairs. The pH change as SO_2 is absorbed is reduced and the buffer anion also serves as an additional basic species, both of which enhance SO_2 absorption. Practical buffers must form soluble calcium salts and have a pk_a between the pk_a of $SO_2 \cdot H_2O$ (≈ 2) and the pk_a of $CO_2 \cdot H_2O$ (≈ 7). For maximum effectiveness the pk_a should be near the mean pH of the liquid boundary layer which is usually in the range of 4.2 to 5.2 (28). From a practical standpoint the buffer must also be inexpensive and stable and nonvolatile under scrubbing conditions. The chemical and practical aspects of buffers in limestone scrubbing are discussed by Hatfield and Potts (78), Rochelle (28), and Rochelle and King (79).

The choice of buffer additives is essentially limited to a number of organic carboxylic acids. Hatfield and Potts investigated over 100 such acids and evaluated glycolic, adipic, benzoic, and phthalic acids in laboratory tests. All were found effective in increasing SO_2 absorption, a conclusion substantiated by small-scale scrubbing tests with benzoic acid. Rochelle and Rochelle and King also provide extensive discussions of the chemistry as well as the practical aspects such as volatility, stability, and cost of organic buffers. They concluded that organic acids other than benzoic, such as acrylic, maleic, adipic, and glycolic, would be more effective and economical.

An extensive evaluation of adipic acid as a buffering additive was undertaken by EPA in 1978, first at the IERL-RTP pilot plant and in June at the Shawnee test facility. Adipic-acid-enhanced limestone scrubbing constituted a major portion of the Shawnee test facility studies from 1978 to 1980. The results of these tests have been reported by Head and others (80), Burbank and others (81), and Wang and Burbank (82).

Initial tests at the IERL-RTP pilot plant confirmed the anticipated effects of adipic acid in improving SO_2 removal efficiency and limestone utilization (80). In addition, it was found to be effective at high chloride levels and to have little effect on forced oxidation. In both limestone and lime scrubbing tests at the Shawnee test facility these results were further quantified (81,82). The initial tests were made with both TCA and venturi - spray tower absorbers, with and without forced oxidation, and at adipic acid concentrations in the range of 1,500 to 3,500 ppm in the slurry. Improved removal efficiencies and absorbent utilization, little effect on forced oxidation, and little adverse effect by chlorides were observed. The control of pH in the absorbers proved more difficult in some cases, however. Also, degradation of the adipic acid, characterized by the odor of valeric acid, that had been observed in the IERL-RTP pilot plant tests, was also evident in the Shawnee tests. As a result, adipic acid consumption was up to five times the theoretical requirements.

Subsequent tests in 1979 and 1980 were conducted with limestone scrubbing using venturi and spray tower absorbers alone and combined, without forced oxidation and with different forced-oxidation schemes (81,82). These tests further quantified the previous results. Adipic acid degradation was found to be greater at high pH's and low limestone utilization. An absorber pH of about 5 was found to be optimum in terms of adipic acid consumption, limestone utilization, and sensitivity of SO_2 removal. Results with a venturi absorber suggested that a venturi alone, with adipic acid enhancement, could be effective for low-sulfur coal applications.

Cocurrently, full-scale tests were initiated at the Springfield (Missouri) City Utilities Southwest Power Plant (83) and on an industrial boiler (84). Large increases (from 50%-60% to about 90%) in SO_2 removal were attained in both cases.

Further tests of adipic acid degradation (85,86) confirmed that pH's below 5 could reduce, but not eliminate, the degradation and that forced oxidation could increase degradation even at low pH's. Appreciable loss was also found to occur because of coprecipitation by adsorption on the calcium sulfite crystals, a condition suggesting that forced oxidation to calcium sulfate could also decrease adipic acid loss.

SCALE FORMATION

The formation of solid deposits in the liquid loop and downstream flue gas equipment has been the most persistent and widespread problem in limestone, and to a lesser extent lime, scrubbing. The problem occurred in the first British FGD systems and was general in early U.S. limestone FGD systems (87). To some extent, control of solid formation is a mechanical problem that has responded to the development of improved design and operating procedures. Simple deposition at the absorber dry-wet interface, in low-velocity sections of the liquid loop, and in equipment downstream from the absorber has been controlled by such features as upstream fly ash removal, presaturators, and improved mist eliminator design and cleaning procedures. To a large extent, however, improved control of solid deposition has evolved from a fuller understanding of chemical aspects, particularly the rates and mechanisms of calcium-sulfur salt crystallization and the effects of pH.

Carbonate scaling is largely a phenomenon of the slurry entrance to the absorber. If the slurry pH is too high (above about 6 and 8, respectively, for limestone and lime scrubbing at the Shawnee test facility), $CaCO_3$ can form on nozzles and adjacent absorber surfaces (27). The condition is alleviated by pH control. Sulfite solids formation is also largely an absorber pH phenomenon that occurs most frequently near the absorber outlet if the pH decrease is insufficient to allow formation of sufficient HSO_3^- to compensate for the increase in the total sulfur species. At the Shawnee test facility an outlet pH of about 5.5 has been found satisfactory (26). Sulfite scaling is also a function of liquid saturation levels and crystal nucleation site density, however. Sulfate scaling, the common gypsum scaling of industrial and natural water systems, is a more complex and serious problem. $CaSO_4$ forms hard, pH-insensitive deposits that must be mechanically removed and which can form rapidly under some conditions.

Both calcium sulfite and calcium sulfate are able to exist in solution at appreciable levels of supersaturation for extended periods, even in the presence of their crystalline solids. At lower levels of supersaturation crystallization occurs primarily at sites on existing crystals. At higher levels of supersaturation nucleation processes increase in importance forming deposits on equipment surfaces. The supersaturation nucleation process has been extensively discussed in FGD literature (30,41). Control of sulfate scaling in FGD systems is largely a combination of various design and operating factors that control the levels and rates of increase of sulfate concentration in the absorber and precipitation mechanisms in various parts of the liquid loop. Additives such as those used in industrial water systems have also proven effective (41).

Sulfate formation is largely a function of the oxygen mass transfer rate and occurs largely in the absorber (27) in designs that exclude extensive air-liquid contact in other portions of the liquid loop. Thus the percentage of SO_2 absorbed, which is subsequently oxidized, can be controlled to some extent by any physical or chemical changes that increase the SO_2 mass transfer rate relative to the O_2 mass transfer rate. Other than design or flue gas compositional changes, this approach is limited to increasing the

SO_2 absorption efficiency by chemical changes such as additives or higher stoichiometries (higher pH's), possibly combined with reduced liquid rates. Higher pH's also reduce oxidation rates by decreasing the quantity of dissolved sulfite available for oxidation. Conversely, the saturation can be maintained below critical levels by increasing the liquid rate, thus decreasing the quantity of SO_2 and O_2 absorbed per volume of liquid, i.e., the "make per pass."

It is also necessary to consider the degree of sulfate saturation of the liquid entering the absorber insofar as it affects the absolute increase in saturation in the absorber. This is largely a function of the hold tank, or tanks, regimen (88). Sufficient time must be allowed in terms of volume and degree of back mixing to obtain the necessary sulfate removal from solution, either as gypsum or enclosed in the sulfite precipitate.

A third factor affecting scaling is the quantity of existing crystal sites, in terms of surface areas, available for precipitation to proceed through crystal growth. Crystal area is largely a function of concentration and size. The area is thus not a direct function of solid $CaSO_4$ concentration although slurry solids is usually a sufficiently accurate measure for scale control (27). Grinding has also been proposed as a further means of control of seed crystal area (30).

Practical sulfate scale control within limits can be attained by various combinations of these factors--oxidation, pH, and seed crystals--depending on the requirements of the system. At the two extremes, control of oxidation below 20% normally allows sufficient sulfate removal with the calcium sulfite while forced oxidation provides sufficient seed crystal area to prevent nucleation.

WASTE TREATMENT AND DISPOSAL

Important advances in waste disposal techniques have paralleled the development of limestone and lime scrubbing FGD, encouraged by the economic and physical problems of dealing with untreated waste sludge as well as the embodiment of environmental concerns in legislation such as the Resources Conservation and Recovery Act of 1976 (RCRA). Among these advances have been a greatly expanded body of information on the chemical and physical characteristics of calcium-based FGD wastes and the development of technologies for treatment such as oxidation, stabilization, and fixation, which have led to improved dewatering and increasing use of landfill disposal in place of ponding.

In the early 1970's ponding disposal was used or planned at 60% of the existing or planned utility FGD facilities (89). In 1980 ponding remained the primary disposal practice for operational units but landfill was the dominate practice for units under construction and was almost solely planned for contracted units (90). In addition, the use of unlined ponds diminished, although ponds are often identified as lined if the in situ soil is considered impervious (91). A trend no less pronounced is the growing use of stabilization and fixation whether for pond or landfill disposal. Utilization is also being considered as a practical means of reducing disposal requirements, at least partially in some cases, by marketing gypsum or use as a construction material (92).

Among the extensive and diverse discussions of waste disposal found in FGD literature, two projects, one by EPA and one by the Electric Power Research Institute (EPRI), have produced a body of open literature that provides a comprehensive view of the field. In 1974, EPA expanded its research and development efforts in FGD waste disposal into an integrated program dealing with power plant waste and water problems. This program, "Control of Waste and Water Pollution from Coal-Fired Power Plants," encompasses a number of programs related to disposal of limestone and lime scrubbing FGD wastes. Results of the program, as well as other related work, have been summarized by The Aerospace Corporation (93). Among these studies are a data base for development of standards and regulations (94), field evaluations at the Shawnee test facility (95), and a series of economic evaluations of disposal methods by TVA (96,97,98). Other general summaries and assessments of FGD waste disposal have also been published (99). EPRI has also produced an integrated series of studies under their projects RP 786 and RP 1260, which culminated in an FGD sludge disposal manual (92) dealing with a broad spectrum of FGD waste disposal technology and practices. The project also produced an assessment of fixation technology (100), an FGD waste data base (101), studies of chemical and physical properties (102,103), and a review of sludge dewatering technology (104).

From the standpoint of this study, waste dewatering techniques and stabilization and disposal practices are the subjects of primary concern. Mechanical dewatering usually consists of a primary dewatering to 30% to 40% solids in thickeners, clarifiers, or ponds. This may be followed by further dewatering to 60% or more solids by vacuum filters or, less commonly, centrifuges.

Conventional circular gravity thickeners with a rake mechanism are the most common primary dewatering devices. Their application to FGD and other waste treatment has been described in EPA (105) and EPRI (92,104) manuals and reports. Normally solids concentrations from 10% to 15% to about 30% to 40% are attained although specific sludge properties have large effects (104) on outlet concentrations.

Hydroclones and high-rate thickeners have also been considered for primary dewatering (92). Neither has been extensively evaluated, however. Brief tests with a high-rate Lamella Settler at the Shawnee test facility (26,27) showed it to be superior in clarifying and equal in thickening to a conventional thickener. Similar comparisons with a hydroclone produced the opposite results; the overflow was high in solids and the underflow contained up to 60% solids. Such factors as erosion and energy requirements for these methods remain to be resolved.

Drum-type vacuum filters are most commonly used for FGD sludge dewatering (92). Belt-type filters are also used; they have advantages in performance and maintenance but are usually more expensive (92). Filters operating with a thickener underflow of high-sulfite sludge can produce a filter cake of 50% to 60% solids depending on the operating conditions and sludge characteristics. Filter cake solids of over 80% are attainable with gypsum sludge.

Centrifuges are less commonly used for secondary dewatering. They generally produce a cake of 60% to 70% solids when fed with high-sulfite thickener underflow sludge (104). Centrifuges are more subject to wear than filters but are capable of producing a uniform cake over a wide range of feed conditions (104).

Dewatered sludges containing less than about 90% gypsum are generally accorded to be unsatisfactory for landfill disposal (26). They are easily reliquefied, have poor bearing strengths, and are highly compressible. Some form of stabilization is essential for landfill disposal in most cases. Stabilization, generally defined as any process that increases the physical stability of the waste, is limited by practical considerations to mixing it with soil or dry fly ash or the use of chemical additives in a fixation process.

Stabilization by addition of dry fly ash has been studied as a potential route to landfill disposal (106) and has been used in full-scale operations (100). Stabilization by fixation, frequently involving dry fly ash as a pozzolan, is more common, however. Numerous commerical fixation processes have been offered by vendors and are well described in FGD literature (100). The most widely used process is the IU Conversion System's Poz-O-Tech process, which was operational at four power plants in 1980 (107). It consists of blending dry fly ash and lime with dewatered sludge to produce, as the fly ash and lime react, a soil-like solid. A similar process, also in full-scale use, is the Dravo Lime Company's process using Calcilox, a processed blast-furnace slag, as the fixative. American Admixtures Company also operates a fly ash and lime fixation process. These and a variety of other processes using Portland cement or a variety of inorganic and organic chemicals are described by Santhanam and others (108) and Duvel and others (100).

Forced oxidation generally produces a waste essentially composed of gypsum. With suitable designs waste composed of nearly pure gypsum can be produced. The crystal morphology that produces the superior dewatering characteristics and bulk density of gypsum, compared with calcium sulfite, has been widely discussed (109,24). Normally gypsum wastes can be dewatered to over 80% solids in conventional equipment or in settling ponds (27). This may permit the disposal method termed stacking, as practiced in the fertilizer industry and at gypsum FGD process test facilities (110). The slurry is pumped onto a diked disposal site where it drains, forming a pile that is used to construct successive rises of containment dikes. Perimeter ditches collect the runoff. Alternately, it is possible that the dewatered gypsum can be trucked to conventional landfills and disposed of in the same manner as suitably stabilized high-sulfite waste.

A substantial body of information on FGD sludge dewatering characteristics has been developed, which permits at least a general projection of dewatering equipment requirements based on the FGD unit design and operating conditions. Such factors as the effects of fly ash, the sulfate/sulfite ratio, type of absorbent, and stoichiometry provide an at least semiqualitative basis for the design of the dewatering system. This information is useful because it permits a more rational basis for the economic evaluation of possible alternate disposal methods. With the wider availability of alternate treatment and disposal methods this assessment acquires additional economic importance.

There still exists, however, a considerable area of uncertainty in such projections of sludge dewatering behavior and equipment performance that requires resolution on a site-specific and vendor-specific level. Tests with sludges from commercial units have shown wide variations in dewatering characteristics not entirely explained by identified physical and chemical differences (104).

ABSORBERS

A variety of absorber types and combinations have been used in limestone and lime scrubbing FGD, a diversity reflecting the numerous approaches to efficient and practical processes. From the inception of FGD, conventional packed absorbers found limited application in limestone and lime scrubbing, primarily because of scaling and plugging, particularly if operated at the low liquid rates their high efficiency permits. Consequently, FGD absorbers have been characterized on one hand by simple internal designs (such as venturis of varied types, columns with open grids or sieve trays and spray absorbers with little internal structure) and on the other hand by proprietary designs.

At the Shawnee test facility these FGD applications were originally represented by a variable-throat venturi combined with a spray tower, a Turbulent Contact Absorber (TCA), and a marble-bed absorber. The marble-bed absorber found application in early commercial utility FGD systems (111), but did not prove satisfactory. Studies at the Shawnee test facility after the early 1970's involved only the venturi - spray tower, in combination and separately, and the TCA. These continue to be representative of current FGD practice, although current trends suggest an increasing reliance on various types and modifications of spray towers and a declining use of venturis and the TCA (90).

Venturi

The venturi scrubber in its many variations has been an industry standard, mostly for particulate collection, since its appearance as an identified type in the 1940's (112). Any absorber operating primarily as an atomizing device on the venturi principle may be regarded as a venturi, including those employing restrictions formed by plate orifices and rods. In its purer form, however, the venturi absorber bears some resemblance to the classical venturi orifice. The operating principle of the venturi involves introduction of a stationary liquid into a rapidly moving gas. Throat velocities range from 200 to 800 ft/sec in various designs. The liquid is shattered into droplets and accelerated to the gas velocity. It provides, during less than a second of droplet acceleration, a large surface area of liquid in chaotic motion and a large number of droplets which collect particles by inertial impact because of the large difference in particle and liquid velocities. The venturi is an efficient particle collection device but a less effective absorber of gases because of the limited contact time. Single-stage venturis in simple limestone scrubbing systems such as the one at the Shawnee test facility typically provide less than 50% SO_2 removal (27). More reactive absorbents (22) and multistage venturis (113) can be used to improve removal efficiency.

The advantages of the venturi lie in its compactness and simplicity of design, which lead to a low initial cost and low maintenance costs. The major disadvantage is the high pressure drop that is inherent in the design. Most of the energy consumed in a venturi absorber is used to accelerate the liquid and is lost as irreversible drag-force work (114).

The venturi has been the subject of considerable study, both in the area of particulate collection and as an SO_2 absorber. Much of this work has dealt with pressure drop and efficiency. Calvert (115) developed expressions relating pressure drop to particulate collection efficiency. Boll (116) developed similar expressions incorporating considerations of duct configurations and wall friction, later expanded by Hollands and Goel (117) and Goel and Hollands (114). Behie and Beeckmans (118) developed models incorporating aerosol collection efficiency that differentiated the effects of downstream separation equipment. Models for gas absorption have also been developed, such as those by Kerr (119) for SO_2 using limestone and lime slurries among others. Uchida and Wen (120) discuss the role of drop formation and heat transfer in developing an SO_2 absorption model.

Spray Towers

The distinction between a spray tower (more precisely a spray column) in which internal structures are used for gas distribution and a packed column containing similar structures to induce changes in gas velocity and liquid flow is not always manifest. In its simplest form, such as the spray tower at the Shawnee test facility, the absorber contains essentially no internal structure other than spray headers.

Spray towers are not widely used in commercial extraction processes, a factor ascribed to their inefficiency because of the excessive backmixing. Modeling based on fundamental principles is less prevalent because of difficulties in evolving general expressions that will be valid for such factors as liquid dispersion and gas flow patterns. Normally no more than two equilibrium stages can be achieved within the limits of height to width imposed by excessive contact of the sprays with the column walls (121).

The primary advantages of the spray column are the simplicity of the internal design and the low gas pressure drop (122). Offsetting these are the high liquid rates and the large energy losses in the spray nozzles. Spray columns also have good turndown capabilities although reduced gas rates may reduce removal efficiency, depending on the effect of gas velocity on drop formation and mixing.

Spray towers used as absorbers typically operate at low pressure drops and large liquid rates of up to 120 gal/kaft^3. The distribution of liquid is highly dependent on such factors as the type, number, and distribution of the nozzles and the liquid nozzle pressure. This allows a considerable latitude of design but complicates scale-up and generalized expressions of performance.

Mobile-Bed Absorbers

Several types of mobile-bed absorbers have found application in FGD, among which the marble-bed absorber and Turbulent Contact Absorber (TCA) represent extremes of the concept. In both cases the packing is an assemblage of spheres that is set into motion by the upward passage of gas, thus providing a packing with less susceptibility to plugging. Some designs evolved as particulate collection devices. In the marble-bed absorber the solid glass spheres move only sluggishly and operate at essentially flooding conditions with an agitated layer of liquid at the top of the bed. In the TCA the bed is greatly expanded and the lightweight spheres are in violent motion.

The TCA has been one of the more successful mobile-bed absorbers applied to FGD. It was originally developed by Dominion Tar and Chemical of Canada and marketed by the Air Correction Division (then Aerotec Industries) of UOP, Inc. (123). The basic design consists of a countercurrent vertical absorber in which the gas moves upward through successive beds of lightweight spheres confined between grids. The sphere beds are placed in turbulent motion by the gas and expand to fill the space between the confining grids. The TCA differs from other mobile-bed absorbers in having a higher gas velocity and a larger free space in the beds. Gas velocities of up to 1,200 ft/min were claimed possible for the original design (123).

The spheres are typically hollow or solid plastic balls in the range of 1.5 inches in diameter that weigh 4 to 7 grams. Problems with wear, cracking, and collapse were encountered in early FGD applications (124). These problems were alleviated at the Shawnee test facility by replacement of hollow polypropylene and other plastic spheres with solid nitrile foam spheres of the same dimensions and mass (27).

The TCA has been extensively investigated to define its operating characteristics in terms of models and basic physical criteria. Uchida and others (125) have identified four regimens of TCA operation: (a) the TCA characteristically exhibits a rapid increase in pressure drop with increasing gas velocity up to the point at which bed fluidization is initiated, (b) thereafter increasing gas velocity has little effect on pressure drop until the bed is fully expanded between the confining grids, (c) at higher gas velocities the spheres congregate against the top grid, and (d) the pressure drop increases rapidly as the gas velocity increases further.

Other studies have investigated wall effect and pressure drop (126), the relationship of pressure drop to bed height (127), and the effects of liquid holdup (128). Models based on data from the Shawnee test facility and the IERL-RTP pilot plant have also been widely discussed (129).

Comparison

The general characteristics of spray tower, venturi, and TCA absorbers have been summarized for limestone FGD applications (130) as shown below:

	Venturi	Spray tower	TCA
SO_2 removal	Low	Good	Good
L/G, gal/ksft3	20-50	70-110	50-85
Gas ΔP, in. H_2O	8-20	1-3	6-12
Gas velocity, ft/sec	125-300	5-25	6-11
Plugging resistance	Excellent	Excellent	Good

The venturi differs considerably in operating characteristics from the TCA and spray tower in its high gas velocity, high pressure drop, and low SO_2 removal efficiency. At equivalent removal efficiencies the TCA and spray tower represent different balances of gas pressure drop and liquid recirculation rate. The higher pressure drop of the more efficient sphere beds of the TCA permits lower L/G ratios, compared with the spray tower. In terms of total power consumption, however, neither approach has a distinct advantage, as shown in published comparisons (130).

	Spray tower	TCA
L/G, gal/kaft3	105	83
ΔP, absorber, in. H_2O	2	10
ΔP, system, in. H_2O	18	26
Power, MW	9.1	9.4

CONSTRUCTION MATERIALS

Corrosion and abrasion, as they affect materials of construction and maintenance, have important bearings on FGD economics. Corrosion, aggrevated by the presence of chlorides, low pH's, the abrasive slurry, and scale, was often unexpectedly severe in early limestone and lime installations (131). Thermal and chemical-property excursions not infrequently added to these problems by increasing corrosion rates and damaging protective coatings. Scale, by creating isolated microzones with extreme chemical conditions, also added to corrosion problems.

Many forms of corrosion occur in FGD systems. Pitting, stress corrosion cracking, crevice corrosion, intergranular corrosion, general corrosion, and erosion-corrosion are all widely reported (132). Areas of particular sensitivity are dry-wet interfaces; areas of high fluid velocity such as throats, nozzles, and pumps; and the duct areas in contact with wet gas between the absorbers and reheaters. Stress corrosion cracking is common in areas subject to stress and exposed to chlorides (133). Stacks have also been subject to rapid deterioration particularly if the clean flue gas is not reheated.

A large body of information has developed on materials of construction for FGD systems although extensive comprehensive compilations have been few. There are numerous general and specific discussions in trade literature (134), publications of the National Association of Corrosion Engineers (135), and in many symposium papers dealing with experimental and full-scale systems (136). Progress reports of the Shawnee test facility discuss the results of a continuing TVA corrosion-tag study as well as the performance of the systems themselves (137). Maiya (138) has summarized many such studies in a general review of materials of construction in FGD systems. Rosenberg and others (139) conducted a survey of construction materials and experience at full-scale utility FGD systems that also deals at length with published materials-related information.

Construction materials range from unprotected carbon steel through numerous alloys and various types of inorganic and organic materials. Usually the components of a particular FGD system represent a spectrum of these materials.

The diversity of construction materials used in different FGD systems for similar functions is a result in part of different design philosophies and the various process differences that exist among sites and among processes. Factors such as fly ash content of the scrubbed gas or the degree of water reuse, as they affect equilibrium levels of scrubbing liquid solids and dissolved impurities, may cause substantial erosion and corrosion differences between otherwise similar processes. There is also a lack of long-term operating experience of the type necessary to define the optimum materials in terms of complex relationships among initial costs, maintenance costs, and reliability.

The use of unlined carbon steel is usually limited to use for ducts in dry-gas areas and for some equipment in the feed preparation and dewatering areas. Lined carbon steel is commonly used for ducts, absorbers, tanks, and piping. Both rubber and filled organic linings are common. Stainless steels that do not contain molybdenum, such as type 304, have not been satisfactory in wet areas. Type 316 and 316L (low carbon) have found wide application in wet gas and absorbent liquid areas although type 316 is subject to rapid corrosion under some conditions of pH and chloride concentration (140). High strength, low alloy (HSLA) steels such as Cor-Ten, which depend on a dense oxidized layer for corrosion resistance, are also widely used, particularly in the flue gas system. Wet conditions, particularly if erosive, may cause rapid corrosion, however (133).

Molybdenum-containing stainless steels are regarded as necessary or desirable in FGD applications where chlorides are present. These, of metallurgical necessity, also contain higher percentages of chromium and nickel. Other stainless steel types such as Carpenter-20, Uddleholm-904L, E-Brite 26-1, and C-Brite 29-4 have also been widely used (141).

Medium alloys containing appreciable nickel such as Carpenter-20, Inconel 825, Hastelloy G, and high alloys containing over 50% nickel, such as Inconel 625 and Hastelloy G, are also used when severe conditions are anticipated (142). These are most frequently used in stressed parts such as reheater tubes and fans.

Inorganic nonmetallic materials fall primarily either into the general class of high-aluminum bricks, shapes, and cements or into the general class of silicon carbide bricks and shapes. They combine good to excellent chemical resistance with high thermal and abrasion resistance. Acid-resistant bricks provide a conventional lining material for ducts and stacks. Alumina bricks and shapes provide more abrasion resistance for linings in such equipment as pumps, venturi throats, and nozzles. Silicon carbide materials provide greater chemical and abrasion resistance in similar applications.

A variety of proprietary high-aluminum hydraulic cements are used as mortars and trowel- or spray-applied linings (139). Organic mortars are also used (132).

The brittleness of inorganic lining material presents a real problem of damage through thermal shrinkage and expansion, vibration, and maintenance activities. Penetration by mortar degradation may also occur. Membrane linings may be used between the substrate and the sheathing to protect the substrate from such penetration (132).

Even more diverse in type and application are the organic materials used primarily as liners in FGD applications. General discussions of these materials are provided by Rosenberg and others (139), Maiya (138), and Javetski (132). A large variety of organic polymeric materials find wide application in FGD systems. Some applications of self-supporting fiber-reinforced plastics are made, such as vanes and vessels, but most organic materials serve as liners for carbon steel. Natural rubber and neoprene, applied in sheets using an adhesive, are widely used as liners. Polyesters and epoxies, frequently with a filler of glass flakes or fibers, are also widely used for the same purpose. They are usually applied as uncured liquids, most often by trowelling or spraying either in the fabricating shop or in the field. Polyesters are superior to most plastics in abrasion resistance and epoxies are superior in bonding properties. Other polymeric organics such as vinyls, urethanes, propylenes, and, less frequently, fluoropolymers are also used.

Organic linings are limited to low service temperatures. Rosenberg and others (139) cite temperature limits ranging from about 180°F (wet) to 250°F (dry) for polyesters and epoxies to 360°F (dry) for vinyls and 400°F (dry) for fluoropolymers.

Organic coatings and self-supporting structures provide additional freedom of design and a potentially economic solution to many FGD corrosion problems. Many applications have met these objectives although there have also been failures of varying degree, particularly in early FGD applications (143). In most cases these failures involved causes unrelated or only remotely related to the corrosion resistance of the material itself.

As both organic materials and coatings, these materials are subject to failure by causes to which metals and inorganics are immune or resistant. Thermal damage, during either operating excursions or maintenance, are possible (140). Organic linings are also more susceptible to erosion than most metals and inorganic materials. As a coating they are also susceptible to debonding, a frequent cause of lining failure (139). Debonding can be the result of a complexity of factors, such as compatibility with the substrate, surface preparation, application, and porosity among which specific causes are difficult to identify (139). While patching is a potentially convenient solution, it is not universally regarded as a permanent repair (144).

These construction materials are offered by scores of vendors in many, often proprietary, variations. Organic materials in particular, with almost unlimited variations in polymers, fillers, and application techniques, mostly proprietary, provide little opportunity for rigid taxonomical treatment. This, combined with the lack of long-term experience with the materials in FGD applications, leads to a diversity of application. Rosenberg and others (139) discuss the lack of generally accessible centralized sources of information on construction materials for FGD and the high degree of reliance on sometimes narrowly specialized vendor information or contracted firms who regard their knowledge as proprietary.

THE SHAWNEE LIME/LIMESTONE SCRUBBING MODEL

The computerized Shawnee lime/limestone scrubbing model used to calculate the design and cost data for this study was developed by Bechtel National, Inc., and TVA to model limestone and lime wet scrubbing. The model consists of various subsystems and subroutines that determine FGD material balances, flow rates, design specifications, capital investments, and revenue requirements for different power unit and coal sulfur conditions. The material balance and flow rate portion of the model was developed by Bechtel using theoretical and empirical mathematical models that are correlated where possible with data from the Shawnee test facility. It also incorporates a modified gas-liquid-solid equilibria model developed by Radian Corporation to determine slurry compositions.

TVA developed design and cost subsystems to provide design and equipment cost data for FGD systems based on these data. Additional TVA-developed subsystems determine the capital investment and annual revenue requirements of the process.

The model has been periodically updated to include new or improved data and process developments in FGD. The basic processes in the current model consist of limestone and lime scrubbing; TCA, spray tower, and venturi absorbers; and pond or landfill disposal. Process options include three modes of forced oxidation, magnesium addition, and adipic acid addition. Several dozen additional input and output options provide further flexibility in use of the model.

The overall model capabilities and its use are described in a users manual published by EPA and TVA (145). Similar briefer descriptions have been presented at EPA industry briefings and symposiums (146-149). The original Radian equilibrium model, the rationale of its development, and background data have also been reported in detail (21) and aspects of the model have been discussed elsewhere (150,151). It consisted of a chemical-composition model based on SO_2 and CO_2 gas-liquid equilibria, liquid equilibria of 20-odd dissolved components, and solid-liquid equilibria of 7 solid components which were used to investigate variations in process rates and stream compositions. In the Shawnee model, as modified by Bechtel (22), it serves as a number of subroutines that calculate slurry compositions.

The specific mathematical treatments of material balances, including SO_2 removal efficiencies, are not fully documented in published works. Descriptions of the mathematical treatment of SO_2 removal in spray tower and TCA absorbers are given in Shawnee test facility reports (25,27).

The absorption of SO_2 into scrubbing liquid approximates the mass transfer situation of absorption followed by a chemical reaction, a circumstance for which no comprehensive theoretical basis exists. Such treatment requires mathematical expressions of turbulent fluid behavior and reaction orders that cannot be rigorously defined. Overall mass transfer models are usually based on modifications of general theoretical treatments that differ in concept but mathematically approach similar conclusions in some cases. Standard references (152) and texts (153) provide discussions and access to the literature.

In practice, the mass transfer functions are reduced by a number of simplifying assumptions based on a knowledge of the system and the likely or probable important and unimportant factors. The mathematical expression at once becomes manageable and specific to the situation, to which it can be further correlated empirically. The development of such expressions is discussed in detail by Wen and Fan (129), Rochelle and King (29), Chang and Rochelle (154), and Wen and others (41), for specific FGD applications.

The Shawnee model expression is simplified by the assumptions that liquid-side resistance controls the absorption rate and that liquid-phase reactions are not limiting (that is, dissolved SO_2 does not significantly affect the absorption rate). Both of these assumptions are supported by experimental results (25,27). The simplified expression for the fraction of

$$SO_2 = 1 - \exp\left[- \phi\, k_L^o az/Hv\right]$$

SO_2 removed, contains an enhancement factor, ϕ, to represent the effects of chemical reaction and a group (consisting of a liquid-side mass transfer k_L^o; interfacial area, a; vertical distance, z; Henry's law constant, H; and gas velocity, v) to represent physical absorption. The enhancement factor contains expressions for pH, effective magnesium, flue gas, SO_2 content, and in some cases chloride concentration. The expression is fitted to Shawnee test facility data for each particular absorber and absorbent combination using eight coefficients. The fitted expressions have standard errors of estimate of about 4%. Pressure drop expressions for the three absorbers were developed by fitting expressions containing pertinent variables to Shawnee test facility data. The development of these expressions is discussed in the Shawnee test facility reports and symposium proceedings cited.

The material balances and other physical and chemical data produced by the model are used to develop design and equipment cost data. Total capital investment and annual revenue requirements are calculated in separate subsystems in the manner described in the premises.

There have also been other models of SO_2 absorption and pressure drop based on Shawnee test facility and IERL-RTP pilot-plant data. Wen and Fan (129) and McMichael and others (155) developed absorption models for spray towers and TCA's based on Shawnee test facility and IERL-RTP pilot-plant data. The TCA model was further developed by Wen and Chang (156) to include the effect of hydrodynamic factors. The models were accurate to about ± 5% for both Shawnee and IERL-RTP data, encouraging credence in their use for scaleup. Kerr (119) developed a model for venturi absorbers in which the effects of gas- and liquid-phase and solids dissolution resistances were differentiated. Wen and Uchida (157) also discuss a venturi model that includes limestone and lime scrubbing.

DESIGN AND ECONOMIC PREMISES

The premises used in this study have been developed by TVA, EPA, and others during similar economic evaluations made since 1967. The design premises are formulated to quantify SO_2 emissions from a representative electric utility coal-fired power plant boiler and to establish efficiencies, process flow rates, and other design and operating conditions to meet environmental regulations. The economic premises are designed to represent the many factors affecting costs. They prescribe procedures for determining capital investment and annual revenue requirements based on regulated utility economics. Results are reported in sufficient detail to permit comparisons with previous economics (1,2) and to permit adjustments based on use of alternative premises. The costs are based on 1982 costs for capital investment and 1984 costs for annual revenue requirements.

The economic evaluations are based on a conceptual design developed from the design premises and engineering data such as flow diagrams, material balances, and equipment costs. The base case is a new 500-MW power unit burning 3.5% sulfur, 16% ash bituminous coal, and complying with the 1979 new source performance standards (NSPS) (158). Case variations in coal sulfur content and power unit size illustrate their effects on the economics of the processes evaluated.

DESIGN PREMISES

Coal

The coals used are three eastern bituminous coals containing 5.0%, 3.5%, and 2.0% sulfur, and one western subbituminous coal containing 0.7% sulfur, all on a dry basis. The assumed properties of these coals are based on composites of many samples representing major coal production areas (159,160). The eastern bituminous coals are assumed to have a heating value of 11,700 Btu/lb and an ash content of 15.1% as fired. The heating value and ash content of the western subbituminous coal are assumed to be 8,200 Btu/lb and 6.3% as fired respectively. The compositions of the coals are shown in Table 1.

Power Plant

The power plant site is assumed to be in the north-central region (Illinois, Indiana, Ohio, Michigan, and Wisconsin). The location represents an area in which coal-fired power plants burning coals of diverse

TABLE 1. COAL COMPOSITIONS

Coal	Sulfur, %	Ash, %	Moisture, %	Heat content, Btu/lb	Ultimate analysis C, %	H, %	O, %	N, %	Cl, %
(As-Fired Basis)									
Eastern bituminous, 5.0% S	4.80	15.10	4.0	11,700	65.2	4.0	5.5	1.3	0.1
Eastern bituminous, 3.5% S	3.36	15.14	4.0	11,700	66.7	3.8	5.6	1.3	0.1
Eastern bituminous, 2.0% S	1.92	15.08	4.0	11,700	67.8	3.7	6.0	1.4	0.1
Western subbituminous, 0.7% S	0.48	6.30	29.3	8,200	49.0	3.5	10.7	0.7	0.02
(Moisture-Free Basis)									
Eastern bituminous, 5.0% S	5.00	15.7			67.9	4.2	5.7	1.4	0.1
Eastern bituminous, 3.5% S	3.50	15.7			69.5	4.0	5.8	1.4	0.1
Eastern bituminous, 2.0% S	2.00	15.7			70.6	3.9	6.3	1.4	0.1
Western subbituminous, 0.7% S	0.68	8.9			69.3	5.0	15.1	1.0	0.02

type and source are situated (161,162). The design is based on standard design practices (163,164) and current trends in utility boiler construction (165,166). The base-case power unit is a new, single 500-MW, balanced-draft, pulverized-coal-fired, dry-bottom boiler. Heat rates are based on power unit size as shown in Table 2. To provide for equitable comparisons, the power units are not derated for energy consumption by the FGD systems evaluated. Instead the energy requirements are charged as independently purchased commodities. Cost estimates are based on a single power unit independent of other units at the site. Power unit size case variations consist of similar 200- and 1,000-MW units. The FGD systems being evaluated are assumed to be installed during construction of the power plant and are assumed to have a 30-year life and to operate at full load for 5,500 hours a year. This operating schedule is equivalent to a total lifetime operation of 165,000 hours.

TABLE 2. POWER UNIT OPERATING TIME

AND HEAT RATE

		New	
Power unit size, MW:	200	500	1,000
Remaining life, years	30	30	30
Full load, hr/yr	5,500	5,500	5,500
Heat rate, Btu/kWh			
Bituminous coal	9,700	9,500	9,200
Subbituminous coal	10,700	10,500	10,200

Flue Gas Compositions

Flue gas compositions are based on combustion of pulverized coal, assuming a total air rate equivalent to 139% of the stoichiometric requirement (the air required for combustion of carbon, hydrogen, and sulfur). This includes 20% excess air to the boiler and 19% additional air leakage to the flue gas in the air heater. It is assumed that 80% of the ash present in each coal is emitted as fly ash. For bituminous coals 92% of the sulfur is emitted as SO_x, while 85% of the sulfur is emitted as SO_x for the western subbituminous coal. The remaining sulfur is removed in the bottom ash and fly ash. No loss of sulfur in the pulverizers is assumed. Three percent of the sulfur emitted as SO_x is SO_3 and the remainder is SO_2. Flue gas compositions and flow rates based on combustion of each of the coals assumed for the study are shown in Tables 3 through 6.

Environmental Regulations

Emissions from new coal-fired utility plants are regulated by NSPS issued under authority of Section 111 of the Clean Air Act as amended. This section

TABLE 3. FLUE GAS COMPOSITION

FOR 5% SULFUR EASTERN BITUMINOUS COAL

(Gas to electrostatic precipitator)

Component	Volume %	Lb-mol/hr	Lb/hr
N_2	75.13	136,900	3,834,000
O_2	5.53	10,080	322,600
CO_2	12.10	22,040	970,000
SO_2	0.30	542	34,750
SO_3	0.01	17	1,343
NO	0.03	59	1,766
NO_2	0.00	3	142
HCl	0.01	12	418
H_2O	6.89	12,550	226,000
	100.00	182,200	5,391,000
Fly ash			50,290
Total			5,441,000

Sft3/min (60°F) = 1,152,000
Aft3/min (300°F) = 1,684,000

Fly ash loading

	Gr/sft3
Wet	5.09
Dry	5.47

Sulfuric acid dewpoint temperature: 316°F

TABLE 4. FLUE GAS COMPOSITION

FOR 3.5% SULFUR EASTERN BITUMINOUS COAL

(Gas to electrostatic precipitator)

Component	Volume %	Lb-mol/hr	Lb/hr
N_2	75.22	137,500	3,852,000
O_2	5.54	10,130	324,200
CO_2	12.33	22,550	992,300
SO_2	0.21	380	24,320
SO_3	0.01	12	940
NO	0.03	59	1,766
NO_2	0.00	3	142
HCl	0.01	12	418
H_2O	6.65	12,160	219,100
	100.00	182,800	5,415,000
Fly ash			50,050
Total			5,465,000

Sft3/min (60°F) = 1,156,000
Aft3/min (300°F) = 1,690,000

Fly ash loading

	Gr/sft3
Wet	5.05
Dry	5.41

Sulfuric acid dewpoint temperature: 308°F

TABLE 5. FLUE GAS COMPOSITION

FOR 2% SULFUR EASTERN BITUMINOUS COAL

(Gas to electrostatic precipitator)

Component	Volume %	Lb-mol/hr	Lb/hr
N_2	75.24	137,800	3,860,000
O_2	5.54	10,100	324,800
CO_2	12.52	22,920	1,009,000
SO_2	0.12	217	13,900
SO_3	0.00	7	537
NO	0.03	59	1,766
NO_2	0.00	3	142
HCl	0.01	12	418
H_2O	6.54	11,970	215,600
	100.00	183,100	5,426,000
Fly ash			49,480
Total			5,475,000

Sft3/min (60°F) = 1,158,000
Aft3/min (300°F) = 1,692,000

Fly ash loading

	Gr/sft3
Wet	4.98
Dry	5.33

Sulfuric acid dewpoint temperature: 297°F

TABLE 6. FLUE GAS COMPOSITION

FOR 0.7% SULFUR WESTERN SUBBITUMINOUS COAL

(Gas to electrostatic precipitator)

Component	Volume %	Lb-mol/hr	Lb/hr
N_2	71.13	156,300	4,377,000
O_2	5.25	11,520	368,700
CO_2	11.89	26,120	1,150,000
SO_2	0.04	79	5,063
SO_3	0.00	2	196
NO	0.02	54	1,627
NO_2	0.00	3	131
HCl	0.00	4	132
H_2O	11.67	25,630	461,800
	100.00	219,700	6,365,000
Fly ash			32,640
Total			6,397,000

Sft3/min (60°F) = 1,389,000
Aft3/min (300°F) = 2,030,000

Fly ash loading

	Gr/sft3
Wet	2.74
Dry	3.10

Sulfuric acid dewpoint temperature: 280°F

requires EPA to set Federal emission limitations that reflect the degree of control that can be achieved by using the best available control technology (BACT). On December 23, 1971 EPA issued NSPS to limit emissions of SO_2, NO_x, and particulate matter from utility power plants (167). In 1979 EPA revised the NSPS (158), which are shown in Table 7. The controlled outlet SO_2 emission and SO_2 removal efficiencies for premise coals are shown in Figure 2 and tabulated in Table 8.

TABLE 7. 1979 REVISED NSPS EMISSION STANDARDS

SO_2

70% SO_2 removal (minimum) to a maximum SO_2 emission of 0.6 lb SO_2/MBtu
0.6 lb SO_2/MBtu maximum emission up to 90% SO_2 removal
90% SO_2 removal (minimum) to a maximum SO_2 emission of 1.2 lb SO_2/MBtu
1.2 lb SO_2/MBtu maximum emission

NO_x

Bituminous coal - 0.6 equivalent lb NO_2/MBtu
Subbituminous coal - 0.5 equivalent lb NO_2/MBtu
Lignite - 0.6 equivalent lb NO_2/MBtu

Particulate

0.03 lb/10^6 MBtu

TABLE 8. PREMISE COAL EMISSION STANDARDS

Coal	Equivalent SO_2 content of coal, lb SO_2/MBtu	Overall equivalent SO_2 removal efficiency, %	Equivalent SO_2 removal required in FGD system, %[a]	Controlled outlet emission, lb SO_2/MBtu
Eastern bit., 5.0% S	8.21	90.0	89.1	0.82
Eastern bit., 3.5% S	5.74	89.6	88.7	0.60
Eastern bit., 2.0% S	3.28	81.7	80.1	0.60
Western subbit., 0.7% S	1.17	70.0	64.7	0.35

a. Based on FGD as the only SO_2 control device and the previously defined sulfur retention in the ash.

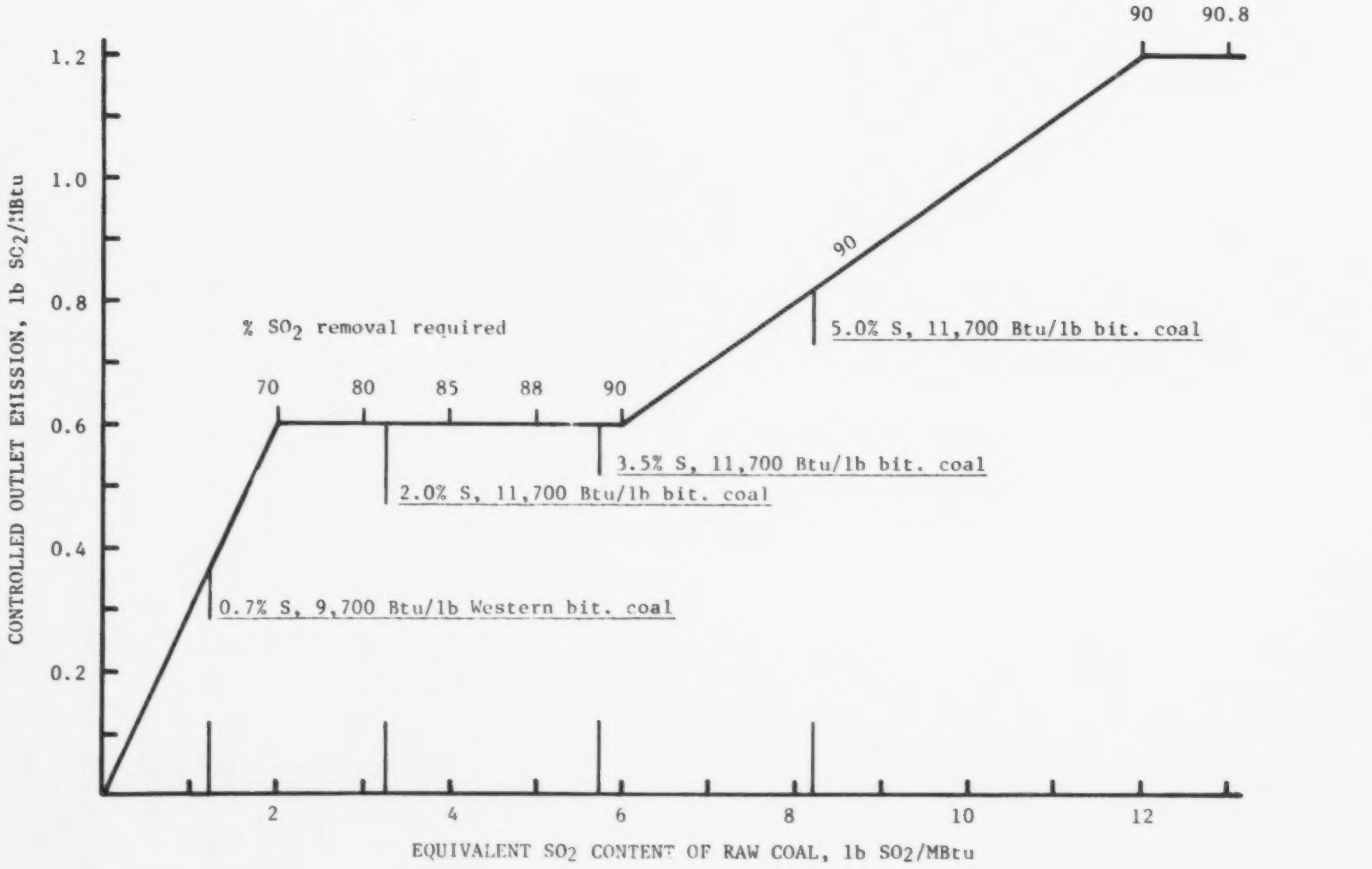


Figure 2. Controlled SO_2 emission requirements for 1979 NSPS. Premise coals, shown underlined, are based on premise boiler conditions.

Particulate Matter

Cold-side (after the air heater) ESP's sized to meet the 0.03 lb/MBtu standard are assumed for particulate matter control. The costs for ash collection and disposal are not included in the economic evaluations.

Flue Gas Desulfurization

The conceptual design of the FGD system meets applicable emission standards and reflects a practical operating approach. FGD systems are close coupled to the power unit by a plenum into which the power plant ID fans discharge. The plenum allows the scrubbing systems to be designed for a different number of trains than the number of power plant ducts (to account for limitations in the available size of absorber units), and it facilitates the use of spare absorber trains. To minimize flow control problems that can result from this design, separate booster fans are provided for each train of the FGD system. The power plant ID fans downstream from ESP's are assumed to discharge into the FGD system plenum as they would into the stack plenum if the FGD system was not installed. The FGD booster fan downstream from the reheater overcomes the pressure drop in the FGD train.

Absorber trains are sized for a maximum of 125 MW of flue gas up to a maximum of 513,000 sft^3/min (60°F). Thus for full scrubbing, the 500-MW base case requires four (125 MW) operating trains and the 200- and 1,000-MW case variations require two (100 MW) and eight (125 MW) operating trains, respectively. It is assumed that the annual availability of an absorber train is 85% and that no scrubbing time is lost during startup. Spare absorber trains are provided as described below. All are closed-loop processes in which the only liquid water loss is in the waste. Material balances, including loss of additives, are determined on this basis.

Emergency Bypass--

Because the 1979 NSPS allow emergency bypass around the FGD system under some conditions if spare scrubbing capacity is provided, spare absorber trains and provision for bypass of 50% of the gas scrubbed are included in all FGD economic evaluations. The 1,000-MW case variation with eight operating absorber trains is provided with two spare trains. Units on smaller boilers are provided with one spare train. An emergency bypass of 50% of the scrubbed gas is assumed to be an economic balance between the higher cost of providing additional bypass and the small likelihood of multiple absorber train failures making higher bypass rates necessary. The bypass is installed as two identical ducts from each end of the inlet plenum to the stack plenum.

Partial Scrubbing--

For the 0.7% and 2.0% sulfur coals, scrubbing a portion of the flue gas at a 90% SO_2 removal efficiency and combining it with the remaining flue gas is more economical than scrubbing all of the flue gas at a lower SO_2 removal efficiency. In such cases the bypassed gas duct requirements and the emergency bypass capability are combined in the same duct. The ducts are sized to handle both the flue gas normally bypassed and the emergency bypass of 50% of the gas normally scrubbed. For the 0.7% sulfur coal cases 28% of the gas is bypassed. For the 2.0% sulfur coal cases 11% of the gas is bypassed at the 90% SO_2 removal efficiency used for the standard process conditions. (Ninety-five percent SO_2 removal efficiencies with higher bypass rates are also evaluated as special additive-enhanced cases.)

Ductwork--

Square ductwork with 2-inch insulation is used for the inlet plenum and absorber trains. To prevent ash settling, a gas velocity of 50 ft/sec is used for the inlet plenum, all ductwork, and the emergency bypass. A gas velocity of 25 ft/sec is used for the reheater section. Duct material is 3/16-inch Cor-Ten steel when the gas temperature is higher than 150°F and 3/16-inch stainless steel when the gas temperature is lower than 150°F.

Removal Efficiencies--

It is assumed that 50% of the SO_3, 95% of the HCl, none of the NO_x, and 50% of the remaining fly ash in the flue gas are removed in the FGD system. It is assumed that 5% of the SO_2 is removed in the presaturator and that the remaining SO_2 removal takes place in the absorber.

Spare Equipment--

Equipment is spared in accordance with general field practice. This consists of a spare train of crushing and grinding equipment, slakers, sludge filters, pumps, and the spare absorber train or trains.

Mist Eliminator--

The mist eliminator is a three-pass, open-vane, chevron-type. The mist eliminator reduces entrained moisture to a maximum level of 0.1% (by weight) of the flue gas. This maximum level is assumed for calculation of the amount of stack gas reheat required.

Stack Gas Reheat--

Indirect steam reheat to 175°F is provided to evaporate entrained water droplets not removed by the mist eliminator. Approximately one-half of the reheater tubes are made of Inconel 625 and approximately one-half of Cor-Ten. Inconel 625 is highly resistant to corrosion and is used for the first bank of tubes, which increases the flue gas temperature to 150°F. The Cor-Ten tubes follow directly after, raising the temperature of the gas at the exit to 175°F. For cases which utilize partial bypass the scrubbed gas is heated to a temperature less than 175°F because of the heat obtained from the bypassed gas. In these cases, the total amount of reheater surface decreases; however, the percentage of Inconel 625 tubes increases to as much as 100% (for reheat to 150°F or less).

Raw Materials--

Raw materials storage capacity is normally 30 days unless process or industry practice differs. Standard raw material characteristics are shown in Table 9.

TABLE 9. RAW MATERIAL CHARACTERISTICS

	Size as received	Ground size	Analysis[a]	Bulk density, lb/ft^3
Limestone	0 x 1-1/2 inch	90% to pass 325 mesh	95% $CaCO_3$ 0.15% MgO 4.85% inerts 5 lb H_2O/100 lb dry limestone	95
	Fineness of grind index factor = 5.7 Hardness of work index factor = 10			
Lime (pebble)	3/4 x 1-1/4 inch	-	95% CaO 0.15% MgO 4.85% inerts 5 lb H_2O/100 lb dry lime	55
MgO	Crystalline powder	-	98% MgO 2% inerts	30
Adipic acid	Crystalline powder	-	99.8% $(CH_2)_4(COOH)_2$ 0.2% inerts	49

a. Limestone and lime analysis on a dry basis. H_2O is based on pounds of dry limestone or lime.

Solids Disposal

The solids generated by the FGD processes evaluated are disposed either by ponding or landfill at a site one mile from the FGD facilities. Sufficient land is provided for disposal for the projected life of the FGD facility. The disposal site is assumed to be an area of low relief with sufficient soil for dike construction or landfill requirements.

Pond--

Disposal ponds are square, earthen-diked enclosures with a median diverter dike. Dikes are constructed from material removed from the impoundment area as shown in Figure 3. The entire impoundment area is lined with 12 inches of clay (assumed available onsite). Pond size and depth are adjusted to minimize the sum of land and construction costs. Pond costs include a 6-foot security fence around the perimeter dike, security lighting, a topsoil storage area, and one upstream and three downstream groundwater monitoring wells.

Landfill--

Square area-type landfills with a 20-foot-high perimeter and a 6-degree cap are used, as shown in Figure 4. After topsoil removal the landfill area is lined with 12 inches of clay (assumed available onsite). A French drain system (perforated pipe surrounded by gravel) and 24 inches of bottom ash are added on top of the clay. The bottom ash layer and French drain system allow the seepage to be collected separately from any runoff and treated for pH adjustment.

Land requirements include the landfill, catchment basin, equipment storage area, topsoil storage area, and a 50-foot perimeter of undisturbed land. Costs for access roads; a 6-foot security fence around the total landfill area; security lighting; and topsoil stripping, replacement, and revegetation are included. One upstream and three downstream groundwater monitoring wells are also included.

ECONOMIC PREMISES

Schedule and Cost Factors

A 3-year construction period, from early 1981 to late 1983, is used. Mid-1982 costs are used for capital investment. Mid-1984 costs are used for annual revenue requirements. These costs represent the midpoint of construction expenditures in 1982 and the midpoint of the first year of operation in 1984. Costs are projected from Chemical Engineering cost indexes (168), as shown in Table 10, using standard estimating techniques (169,170). Frequently used costs are shown in Table 11.

Capital Cost Estimates

The capital investment estimates are divided into three major sections: direct investment, indirect investment, and other capital investment. The sum of direct and indirect investments is the fixed investment.

Direct Investment--

Direct investment consists of the installed costs of all process equipment, including provisions for services, utilities, and miscellaneous; and waste disposal investment. Installation costs are estimated by major

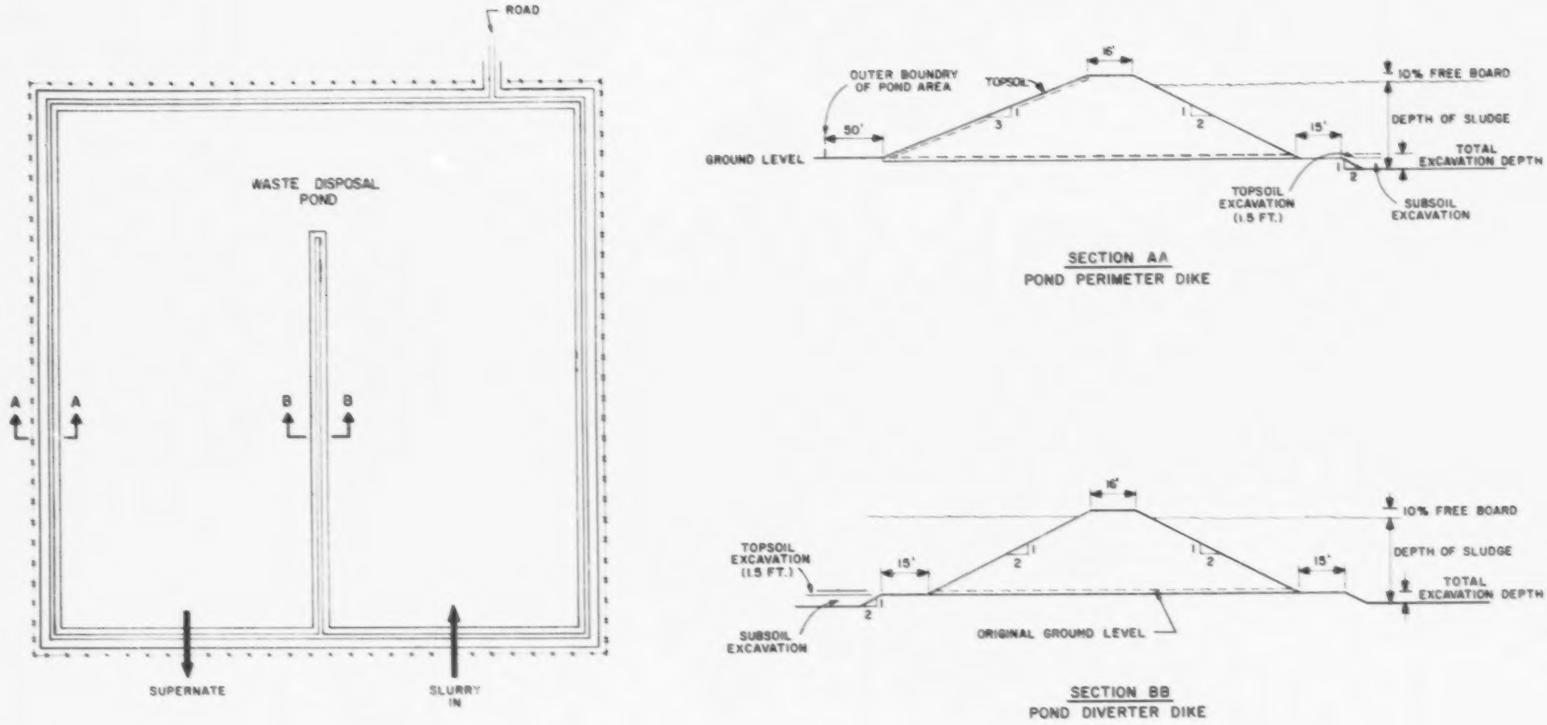

ROAD
WASTE DISPOSAL POND
A
A
B
B
SUPERNATE
SLURRY IN
OUTER BOUNDRY OF POND AREA
TOPSOIL
16'
50'
GROUND LEVEL
3
1
1
2
15'
1
2
10% FREE BOARD
DEPTH OF SLUDGE
TOTAL EXCAVATION DEPTH
TOPSOIL EXCAVATION (1.5 FT.)
SUBSOIL EXCAVATION
SECTION AA
POND PERIMETER DIKE
16'
TOPSOIL EXCAVATION (1.5 FT.)
SUBSOIL EXCAVATION
15'
2
1
1
2
15'
ORIGINAL GROUND LEVEL
10% FREE BOARD
DEPTH OF SLUDGE
TOTAL EXCAVATION DEPTH
SECTION BB
POND DIVERTER DIKE

Figure 3. Pond plan and dike construction details.

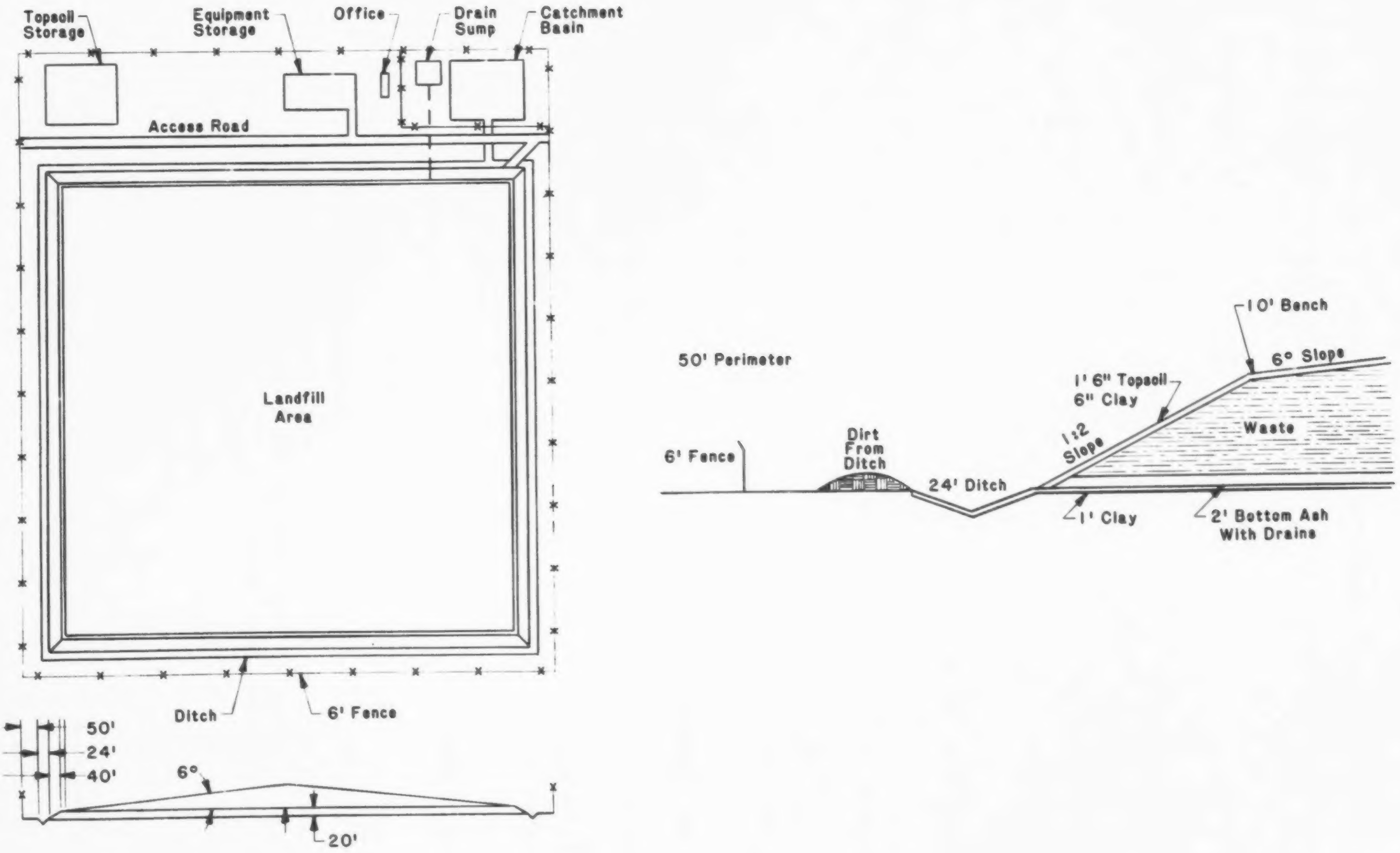

Topsoil Storage
Equipment Storage
Office
Drain Sump
Catchment Basin
Access Road
Landfill Area
Ditch
6' Fence
50'
24'
40'
6°
20'
50' Perimeter
6' Fence
Dirt From Ditch
24' Ditch
1:2 Slope
1' 6" Topsoil
6" Clay
10' Bench
6° Slope
Waste
1' Clay
2' Bottom Ash With Drains

Figure 4. Landfill plan and construction details.

TABLE 11. COST FACTORS

Project Timing	
Start	January 1981
End	December 1983
Midpoint	Mid-1982
First-year operation	1984
1984 Utility Costs	
Electricity	\$0.037/kWh
Steam	\$2.50/klb, \$3.30/MBtu
Diesel fuel[a]	\$1.60/gal
Filtered river water	\$0.14/kgal
1984 Labor Costs	
FGD operating labor	\$15.00/man-hr
Waste disposal labor	\$21.00/man-hr
Analyses labor	\$21.00/man-hr
1984 Raw Material Costs	
Limestone	\$8.50/ton (95% $CaCO_3$, dry basis)
Lime	\$75/ton (pebble, 95% CaO, dry basis)
Adipic acid	\$1,500/ton
Dibasic acid	\$640/ton (dry basis)
Magnesium oxide	\$460/ton
1982 Land Cost	
	\$5,000/acre

These cost factors are based on a north-central plant location.

---

[a] Cost is based on wholesale price of barge-load quantities. Road taxes are not included.

processing area and include charges for all piping, foundations, excavations, structural steel, electrical equipment, instruments, ductwork (all flue gas ductwork is included in gas handling area), paint, buildings, taxes, freight, and a premium for 7% overtime construction labor.

TABLE 10. COST INDEXES AND PROJECTIONS

Year:	1978	1979	1980[a]	1981[a]	1982[a]	1983[a]	1984[a]
Plant	218.8	238.7	257.8	277.1	297.9	320.2	342.6
Material[b]	240.6	264.4	288.2	311.2	336.1	363.0	388.4
Labor[c]	185.9	194.9	210.5	227.3	245.5	265.2	283.7

a. TVA projections.
b. Same as "equipment, machinery, supports" Chemical Engineering index.
c. Same as "construction labor" Chemical Engineering index.

Service facilities such as maintenance shops, stores, communications, security, offices, and road and railroad facilities are estimated or allocated on the basis of process requirements. Included in the utilities investment are necessary electrical substations, process water, fire and service water, instrument air, chilled water, inert gas, and compressed air facilities. Services, utilities, and miscellaneous are estimated as 6% of the total process capital.

All equipment and direct construction costs associated with waste disposal are included in waste disposal costs. For ponds, this includes pond construction costs. For landfills, mobile equipment and construction costs are included. All mobile equipment involved in loading and transporting the waste from the in-process storage area, as well as working the landfill, is included in landfill equipment. The landfill construction cost is listed separately from the landfill equipment. The sum of total process capital, services, utilities, and the waste disposal cost is the total direct investment.

Indirect Investment--

Indirect capital costs cover fees for engineering design and supervision, architect and engineering contractor, construction expense, contractor fees, and contingency. Listed in Table 12 are the percentages used to calculate these costs.

Other Capital Investment--

The allowance for startup and modifications is applied as a percentage of the total fixed investment. Since the startup and modification costs for the waste disposal area are assumed to be negligible, this allowance is calculated using only the process fixed investment.

TABLE 12. INDIRECT CAPITAL COST FACTORS

	Indirect investment, % of direct investment				
	Process			Waste disposal	
	1,000 MW	500 MW	200 MW	Pond	Landfill
Engineering design and supervision	6	7	8	2	2
Architectural and engineering contractor	1	2	3	1	1
Construction expense	14	16	18	8	8
Contractor fees	4	5	6	5	5
	25	30	35	16	16
	Contingency, % of direct investment plus other indirect investments				
	10	10	10	10	20
	Allowance for startup and modification, % of fixed investment				
	8	8	8	0	0

The cost of borrowed funds (interest) during construction is 15.6% of the fixed investment (both process and waste disposal). This factor is based on an assumed 3-year construction schedule and is calculated with a 10% weighted cost of capital, with 25% of the construction expenditures in the first year, 50% in the second year, and 25% in the third year of the project. Expenditures in a given year are assumed uniform over that year. Startup costs are assumed to occur late in the project schedule so that there are no charges for the use of money to pay startup costs. No royalty charges are included because the lime and limestone processes are generic.

All land associated with the process and waste disposal area is charged to the process. The cost of land is $5,000 per acre.

Working capital is the total amount of money invested in raw materials, supplies, finished products, accounts receivable, and money on deposit for payment of operating expenses. Working capital is the equivalent cost of 1 month's raw material cost, 1.5 months' conversion cost, 1.5 months' plant and administrative overhead costs (all of the above are shown on the annual revenue requirements sheet), and 3% of the total direct capital investment. One month is defined as 1/12 of annual costs.

Annual Revenue Requirements

Annual revenue requirements account for recovery of various direct and indirect operating and maintenance costs and capital charges. Annual revenue requirements normally vary from year to year as operating and maintenance costs change and capital charges decline. Thus no single year is necessarily representative of the lifetime costs, and single-year undistorted comparisons cannot be made among processes with different ratios of operating costs to capital charges. In addition, it is necessary to take into account the effect of time on the value of money (i.e., for inflation, the future earning power of money spent, and other factors).

Frequently these factors are accounted for by levelizing (171). Levelization converts all the varying annual revenue requirements to a constant annual value, such that the sum of the present worths of the levelized annual revenue requirements equals the sum of the present worths of the actual annual revenue requirements. The levelized value is calculated by multiplying the revenue requirements for each year by the appropriate present worth factor and summing the present worth values. Then the single present worth value is converted to equal annual values by multiplying the result by the capital recovery factor.

In these premises the operating and maintenance costs are levelized by multiplying the first-year operating and maintenance cost by a levelizing factor. The levelized capital charges are determined by levelizing the percentage of capital investment applied yearly as capital charges. The levelizing factor includes a discount factor reflecting the time value of money and an inflation factor reflecting the effects of inflation during the operating life of the system. The annual discount rate used is 10% and the annual inflation rate used is 6%.

Operating and Maintenance Costs--

Operating and maintenance costs consist of direct costs for raw materials and conversion and indirect costs for overheads. Conversion costs consist of operating labor and supervision, utilities, maintenance, and analysis costs.

Raw materials include consumables required for their chemical or physical properties, other than fuel for the production of heat. In this evaluation the raw materials consist of limestone, lime, magnesium oxide, adipic acid, and dibasic acid (DBA). Raw material costs are determined from vendor quotations or published sources. All costs are delivered costs.

Operating labor and supervision consists of all labor requirements for operation of the FGD-related equipment and waste disposal facility. The allocation of operating labor and supervision man-hours for the FGD system depends on the process complexity, number of process areas, labor intensity of the process, and operating experience. Waste disposal operating labor and supervision depends on the number of pieces of equipment required to operate the landfill. For ponds, FGD operating labor is assumed to include sufficient man-hours to periodically monitor the pond area.

Services such as steam, electricity, process water, and diesel fuel are charged under the utilities heading. Costs for steam and electricity are based on the assumption that the required energy is purchased from another source. This simplifying assumption eliminates the need to derate the utility plant. Process water requirements are defined as any water used by the process being evaluated and are determined from the material balance. Steam requirements are for stack gas reheat. Electrical power requirements are determined from the installed horsepower of operating electrical equipment (excluding the horsepower of spared equipment). Each motor in operation is assumed to be operating at rated capacity although this results in higher power consumptions than would actually occur. Electrical requirements are obtained from the equipment list where the motor horsepower is identified, plus an additional amount for functions such as lighting. Diesel fuel is calculated based on the equipment required for the landfill.

Process maintenance costs are calculated as a percentage of direct process investment which varies with the power unit size. For 200-MW cases the percentage is 9%; for 500-MW cases it is 8%; and for 1,000-MW cases it is 7%. Waste disposal maintenance costs are 3% of the waste disposal direct investment. Analysis costs are based on process complexity and are listed as a single entry.

Plant and administrative overheads include plant services, general engineering (excluding maintenance), interplant communications and transportation, and the expenses connected with management activities. Plant and administrative overheads are 60% of the conversion costs less utilities.

Capital Charges--

Capital charges are those costs incurred by construction of the facility that must be recovered during its life. They consist of returns on equity and debt (discount rate), depreciation, income taxes, and other costs such as insurance and local taxes. In keeping with common practice for investor-owned utilities, the weighted cost of capital is used as the discount rate (172). Depreciation is stated as a sinking fund factor to simplify calculations. An allowance for interim replacement is included. Credits are also included for tax preference allowances. The capital charges are shown below.

	% of total capital investment
Weighted cost of capital	10.00
Depreciation (sinking fund factor)	0.61
Annual interim replacement	0.56
Levelized accelerated tax depreciation	-1.36
Levelized investment tax credit	-1.93
Levelized income tax	4.31
Insurance and property taxes	2.50
Levelized annual capital charge	14.7

The capital charges are applied as a percentage of the total capital investment, including land and working capital. Although land and most of working capital cannot be depreciated and are not subject to investment tax credit, their inclusion has an insignificant effect on capital charges.

The capital structure is assumed to be 35% common stock, 15% preferred stock, and 50% long-term debt. The cost of capital is assumed to be 11.4% for common stock, 10.0% for preferred stock, and 9.0% for long-term debt. The weighted cost of capital is 10.0%. The discount rate is equal to the weighted cost of capital. Other economic factors used in financial calculations are a 10% investment tax credit rate, 50% State plus Federal income taxes, 2.5% property tax and insurance, and an annual inflation rate of 6%. Salvage value is assumed to be less than 10% and equal to removal cost.

The sinking fund factor method of depreciation is used since it is equivalent to straight-line depreciation levelized for the economic life of the facility using the weighted cost of capital. The use of the sinking fund factor does not suggest that regulated utilities commonly use sinking fund depreciation. An annual interim replacement (retirement dispersion) allowance (173) is also included as an adjustment to the depreciation account to ensure that the initial investment will be recovered within the actual rather than the forecasted life of the facility. Tax preference allowances are incentives designed to encourage investment as a stimulus to the overall economy. The basic accounting method used is the flow-through method which passes the tax advantages to revenue requirements as soon as they occur.

Levelized Operating and Maintenance Costs--

Assuming a constant inflation rate, the levelized operating and maintenance costs are determined by multiplying the first-year operating and maintenance costs by an appropriate levelizing factor, L_f. The levelizing factor is calculated as follows:

$$L_f = CRF_e\ (K + K^2 + K^3 + \cdots + K^N)$$

$$= CRF_e\ [K(1 - K^N)]/(1 - K)$$

where: CRF_e = capital recovery factor for the economic life

$K = (1 + i)/(1 + r)$; present worth of an inflationary value

i = inflation rate

r = discount rate

N = book life in years

An inflation rate of 6% ($i = 0.06$) and a discount rate of 10% ($d = 0.10$) are used for new units. The first-year operating and maintenance costs are multiplied by the levelizing factor to obtain the levelized operating and maintenance costs.

SYSTEMS ESTIMATED

In this study, economics are projected for 40 limestone and 12 lime scrubbing processes assessing combinations of (1) limestone or lime absorbent; (2) TCA, spray tower, or venturi - spray tower absorber; (3) onsite pond disposal of sludge without forced oxidation or forced oxidation with landfill; and (4) processes without an additive and processes with adipic acid, dibasic acid, or magnesium oxide additive. The alternatives of full versus partial scrubbing are also investigated. In addition, selected combinations are evaluated for 200-, 500-, and 1,000-MW power units and 0.7%, 2.0%, 3.5%, and 5.0% sulfur coals. The 0.7% sulfur coal is a western subbituminous coal, while the other coals are eastern bituminous coals, as described in the design premises. The combinations of process variables evaluated are shown in Table 13 while the cases evaluated are shown in Table 14. The base-case processes identified in Table 13, which serve as a basis of comparison with other variations, are described in more detail later in this chapter.

Since all of the processes evaluated are variations of limestone and lime scrubbing processes, all bear considerable resemblance in design and function. In terms of the processing areas into which the economic results are divided, the processing equipment is similar and differs primarily in size. Description of the process variations evaluated is facilitated by a description of each of these processing areas, followed by detailed descriptions of four representative basic systems.

MAJOR PROCESSING AREAS

All of the processes are divided into the following process areas: (1) materials handling, (2) feed preparation, (3) gas handling, (4) SO_2 absorption, (5) oxidation (when used), (6) reheat, (7) solids separation (when used), and (8) waste disposal.

Materials Handling Area

The materials handling area consists of equipment to unload 0 x 1-1/2 inch limestone from a railcar or truck and transfer it to a 30-day storage pile with built-in live storage hoppers. The major equipment consists of unloading conveyors, feed conveyors, bucket elevators, a dust collecting system, and feed bins. A scraper tractor is also included to manage the area. Normally the unloading system has a capacity of up to 250 tons/hr and the feed system to the preparation area has a capacity of up to 100 tons/hr.

TABLE 13. PROCESS COMBINATIONS

Limestone Processes

*TCA - sulfite sludge ponding
Spray tower - sulfite sludge ponding
Venturi - spray tower - sulfite sludge ponding
TCA - hold tank forced oxidation - landfill disposal
*Spray tower - hold tank forced oxidation - landfill disposal
*Venturi - spray tower - venturi loop forced oxidation - landfill disposal
TCA - hold tank forced oxidation - adipic acid addition - landfill disposal
Spray tower - hold tank forced oxidation - adipic acid addition - landfill disposal
Venturi - spray tower - venturi hold tank forced oxidation - adipic acid addition - landfill disposal
TCA - bleedstream forced oxidation - magnesium oxide addition - landfill disposal
Spray tower - bleedstream forced oxidation - magnesium oxide addition - landfill disposal
Venturi - spray tower - bleedstream forced oxidation - magnesium oxide addition - landfill disposal
Spray tower - adipic acid addition - sulfite sludge ponding
Spray tower - magnesium oxide addition - sulfite sludge ponding

Lime Processes

TCA - sulfite sludge ponding
Spray tower - sulfite sludge ponding
Venturi - spray tower - sulfite sludge ponding
TCA - hold tank forced oxidation - landfill disposal
Spray tower - hold tank forced oxidation - landfill disposal
Venturi - spray tower - venturi hold tank forced oxidation - landfill disposal
TCA - bleedstream forced oxidation - magnesium oxide addition - landfill disposal
Spray tower - bleedstream forced oxidation - magnesium oxide addition - landfill disposal
*Venturi - spray tower - bleedstream forced oxidation - magnesium oxide addition - landfill disposal

---

*Basic systems.

TABLE 14. PROCESSES EVALUATED

Case variations	TCA-P	ST-P	V-ST-P	TCA-ILFO-L	ST-ILFO-L	V-ST-VTFO-L	TCA-AA-ILFO-L	ST-AA-ILFO-L	V-ST-AA-VTFO-L	TCA-MgO-BSFO-L	ST-MgO-BSFO-L	V-ST-MgO-VTFO-L	ST-AA-P	ST-MgO-P
New plant 200 MW														
0.7% S[a]					x									
2.0% S					x									
3.5% S		x			x			x						
5.0% S					x									
500 MW														
0.7% S[a]		x[d]			x[ce]			x[e]			x			
2.0% S		x[d]			x[ce]			x[e]			x			
3.5% S	x[d]	x[d]	x[d]	x[d]	x[bdeg]	x[d]	x	x[ef]	x	x[d]	x[d]	x[d]	x	x
5.0% S		x[d]			x			x			x			
1000 MW														
0.7% S[a]					x									
2.0% S					x									
3.5% S		x			x			x						
5.0% S					x									

PROCESS ABBREVIATIONS:

Scrubbing alternatives

V - Venturi
ST - Spray tower
TCA - Mobile bed absorber
V-ST - Venturi - spray tower

Sludge treatment and disposal alternatives

P - Ponding
L - Landfill
ILFO - In-loop forced oxidation
BSFO - Bleedstream forced oxidation
VTFO - Venturi hold tank forced oxidation

Additive alternatives

AA - Adipic acid
MgO - Magnesium oxide

---

a. Forced oxidation is not applied in these cases, but 95% natural oxidation is assumed.
b. Base case process.
c. Includes cases of full scrubbing and partial scrubbing.
d. Limestone and lime.
e. SO_2 removal efficiencies of NSPS and 95%.
f. Includes a dibasic acid case.
g. Includes a no reheat case.

For FGD systems designed for lime instead of limestone the material handling area consists of equipment to unload pebble lime from a railcar or truck and transfer it by enclosed conveyor to concrete storage silos at the FGD unit. The lime conveying and handling system is designed for 100 tons/hr. The storage silos have a 30-day storage capacity. Other major equipment in the lime handling area consists of bucket elevators for filling the storage silos and preparation area feed bins, conveyors for moving the lime from the storage silos to the feed preparation area, and a feed bin with an 8-hour capacity.

Feed Preparation Area

The feed preparation area for the limestone systems consists of crushers, wet ball mills, mill product tanks with agitators and pumps, a slurry feed tank with an agitator and pumps, and a dust collection system. The feed preparation unit is designed to produce a 60% solids slurry for addition to the FGD system. The crushers and ball mills are situated in the limestone storage area. The limestone is first crushed to 0 x 3/4 inch in two parallel 50% capacity gyratory crushers and then wet ground as a 60% solids slurry to 90% minus 325 mesh. The minimum mill size used is 100 hp and the maximum size is 2,500 hp. The number of ball mills depends on the limestone feed rate. Generally two operating mills are used and one spare mill is always provided. The slurry is pumped to the slurry feed tank which is located at the FGD unit. This tank has an 8-hour capacity and is equipped with pumps for feeding the slurry to the absorber hold tanks. For systems that use adipic acid, DBA, or magnesium oxide additives, additional equipment is included in the preparation area for their addition. The additional equipment for adipic acid and magnesium oxide includes a 30-day storage silo, feed bins, and metering conveyors that feed the additive into the mill product tank. DBA, which is purchased as a 60% solution, is diluted to a 20% solution and stored in a 30-day-capacity, heated storage tank. Other equipment includes an agitator and a feed pump. The equipment for additives is relatively small and special facilities for receiving these materials are not required.

The feed preparation area for lime systems consists of lime metering conveyors, parallel (the number depends on the flow rate) dual-compartment overflow-type lime slakers, slaker receiving tanks, agitators and pumps, and a single product slurry feed tank with an agitator and feed pumps. The lime preparation area supplies a 20% solids slurry for addition to the absorber hold tanks.

Gas Handling Area

The gas handling area consists of a feed plenum that distributes the flue gas to the individual absorber trains, the absorber ductwork between the feed plenum and the stack plenum, two emergency bypass ducts (one from each end of the inlet plenum to each end of the stack plenum), and one ID fan per absorber train.

The ductwork upstream of the absorbers is constructed of Cor-Ten steel. The ductwork downstream of the absorbers is constructed of 316 stainless steel if the gas temperature is less than 150°F. If the gas temperature is over 150°F (the ductwork between the reheaters and the stack plenum when reheat above 150°F is required) the ductwork is constructed of Cor-Ten steel. Costs for the stack plenum are not included in the FGD system costs. The ID fans are made of Inconel 625 to provide corrosion protection. The two bypass ducts are constructed of Cor-Ten steel and are designed to handle 50% of the scrubbed gas (25% per duct). If some of the flue gas is bypassed, the size of the bypass ducts is increased to include the bypassed gas. For the adipic-acid-enhanced processes and the 0.7% sulfur coal cases, flue gas bypass is used. In all other coal cases, full scrubbing is used.

SO_2 Absorption Area

TCA, spray tower, or venturi - spray tower absorbers are used. The TCA and the spray tower both have a presaturator section into which scrubbing liquid at an L/G ratio of 4 gal/kaft^3 is sprayed to cool and humidify the gas. Typical operating conditions for the absorbers for the various coal sulfur cases evaluated are shown in Table 15. For the adipic-acid-enhanced

TABLE 15. ABSORBER OPERATING CONDITIONS

	TCA	Spray tower				Venturi - spray tower
Coal sulfur, %:	3.5	0.7	2.0	3.5	5.0	3.5
Limestone, No Additives						
L/G, gal/kaft^3	58	115	69	106	109	20/103
ΔP^a, in. H_2O	15.2	7.2	7.0	7.9	7.9	18.0
Bypass, %	0	28	0	0	0	0
Limestone, Adipic Acid						
L/G, gal/kaft^3	45	80	80	80	80	20/50
ΔP, in. H_2O	10.0	6.2	7.0	7.1	7.1	16.6
Bypass, %	7	32	16	7	6	7
Lime, No Additives						
L/G, gal/kaft^3	50	60	60	70	70	20/50
ΔP, in. H_2O	14.0	5.9	6.7	7.0	7.0	16.8
Bypass, %	0	28	0	0	0	0

a. Total system pressure drop.

processes, an adipic acid concentration of 1,500 ppm in the absorber loop is used. This results in an SO_2 removal efficiency of 95% in the absorber and thus allows substantial amounts of flue gas bypass in the lower sulfur coal cases and lesser amounts for the higher sulfur coal cases.

A horizontal, three-pass, open-vane, fiberglass, chevron-type mist eliminator above the absorbers is provided to reduce the entrained moisture content of the scrubbed gas to a maximum level of 0.1% (by weight) of the flue gas. The mist eliminator is washed on a continuous basis on the underside and on an intermittent basis on the topside with fresh makeup water.

The TCA is constructed of neoprene-lined carbon steel and contains four 316L stainless steel grids spaced 4 feet apart, forming 3 absorber beds. Each absorber bed is packed with nominal 1.5-inch-diameter, 6.5-gram, nitrile foam spheres to a static depth of 5 inches. A single bank of sprays above the top bed distributes the absorbent liquid in the absorber. The gas velocity is 12.5 ft/sec.

The spray tower unit is also constructed of neoprene-lined carbon steel. Three 316L stainless steel grids similar to those in the TCA control gas distribution. Four banks of sprays are used for absorbent liquid distribution, one above each of the top three grids spraying downward, and one below the bottom grid spraying upward. The gas velocity for the spray tower is 10 ft/sec.

The venturi - spray tower absorber consists of a venturi scrubber and a spray tower in series, each with a separate absorbent recirculation system. The design and construction of the spray tower is the same as the single spray tower absorber except for piping sizes, which are somewhat smaller because a lower L/G ratio is used.

The venturi is constructed of 316 stainless steel and has an acid-brick-lined throat with a movable axial plug. Flue gas enters horizontally into the top of the vertically mounted venturi and flows downward through the throat. Absorbent liquid is injected into the area above the throat through a peripheral annulus and an axial spray. The venturi is mounted over a catch basin that drains to the hold tank. Flue gas leaves the catch basin horizontally and enters the bottom of the spray tower. The venturi serves as the presaturator and provides some SO_2 removal capability.

Also included in the SO_2 absorption area is the absorbent liquid recirculation system consisting of tanks, piping, and pumps. This equipment varies, depending on whether forced oxidation is used, the type of forced oxidation, and the type of additive.

For processes without forced oxidation, both with and without additives, all three types of absorbers are provided with a single 8-minute-capacity hold tank beneath the absorber into which the absorbent liquid drains by gravity and from which it is recirculated to the absorber. For the processes using a venturi - spray tower without forced oxidation a single tank receives the absorbent liquid from both units and supplies both units.

For processes using the TCA or spray tower with forced oxidation, both with and without adipic acid additive, the tank beneath the absorber has a 5-minute capacity and serves as the oxidation tank (charged to oxidation area costs) and a separate 8-minute-capacity hold tank is provided. For processes using the venturi - spray tower with forced oxidation, both with and without adipic acid, an 8-minute-capacity hold tank for the spray tower and a 5-minute capacity oxidation tank (charged to oxidation area costs) for the venturi are provided. For processes with forced oxidation and MgO additive, bleedstream oxidation is used and the hold tank configuration is the same as the configuration for processes without forced oxidation. All hold tanks are baffled and agitated and made of carbon steel with glass-flake-filled organic polymer linings. Pumps and pipes are rubber-lined carbon steel. A minimum of two pumps and a spare are provided for each function. All pumps are fixed capacity.

Oxidation Area

The oxidation area is included only when forced oxidation is used. (For all 0.7% sulfur coal cases, 95% natural oxidation is assumed to occur and thus no forced-oxidation equipment is necessary.) Major equipment in the oxidation area consists of low-pressure air compressors to supply air to the oxidation tank, the oxidation tank itself, a sparging ring in the base of the tank, an agitator to maintain solids dispersion and improve air distribution, and a bleed pump to transfer the oxidized slurry to the solids separation area. In most cases the oxidation tank is situated directly under the SO_2 absorber. For the venturi - spray tower system, however, the oxidation tank is the venturi hold tank. For the bleedstream oxidation cases the oxidation tank is situated adjacent to the absorber hold tank. When absorber-loop or bleedstream forced oxidation is used, the oxidation area costs include the extra ductwork needed to span the additional tank. Construction materials are the same as those used for the hold tanks and absorbent liquid recirculation equipment.

Reheat Area

All FGD systems are designed for a flue gas temperature of 175°F at the entrance to the stack. Three sources of energy provide this temperature: (1) indirect inline steam reheat, (2) reheat obtained by compression of the flue gas in passing through the ID fans, and (3) reheat obtained by combining the scrubbed gas with gas that has bypassed the scrubbing trains in partial scrubbing cases. The amount of reheat provided by inline steam reheat is calculated by determining the total reheat required and subtracting the quantity of reheat available from compression and bypassed gas. To provide corrosion protection the reheater tubes in contact with the gas up to a temperature of 150°F are constructed of Inconel 625. The remaining tubes are made of Cor-Ten steel. Retractable soot blowers are included with the reheater to keep the tube bundles clean. When using the bypass gas for reheat the bypassed gas is mixed with the wet scrubbed gas in the stack plenum.

Solids Separation and Disposal Areas

Two types of waste disposal are used in this study: direct ponding and dewatering in a thickener and vacuum filters followed by landfill. In both cases the final solids disposal area is assumed to be one mile from the plant.

In the case of ponding, the waste stream (8% solids) is pumped to the pond and pond supernate is returned to the scrubbing area. Major equipment requirements for the ponding cases consist of the pond feed tank and agitator, variable speed disposal pumps, and supernate return pumps. A diagram of the pond area is shown in the premise section.

In the landfill cases the waste stream (8% solids) is pumped to a thickener and then to vacuum filters. Thickener overflow and filtrate is returned to the absorption area. The filter cake (85% solids) is transported to the landfill in trucks. The major equipment for landfill consists of a thickener feed tank and agitator, the thickener, a thickener overflow tank, a filter feed tank with an agitator, rotary drum filters (2 operating and 1 spare in most cases), filter cake conveyors, a filtrate surge tank, a gypsum storage area retaining wall, and mobile equipment. A flow diagram of the solids separation area is shown in Figure 5. A detailed description of the landfill disposal area is given in the premise section.

Configurations

All of the systems evaluated in this study consist of various composites of the areas described above. For descriptive purposes the materials handling and feed preparation areas differ only between the limestone and lime processes. The description of the gas handling area is the same for all processes except for size and the use of bypass in some cases.

One of the major areas of difference between processes is the SO_2 absorption area. The three absorber types combine with two hold tank configurations depending on the use of forced oxidation and the type of forced oxidation. The hold tank configurations are:

- A single hold tank per train under the absorber for all processes without forced oxidation, all processes with bleedstream forced oxidation, and all 0.7% sulfur coal (natural oxidation) cases.
- A single hold tank per train downstream of the oxidation tank for all TCA and spray tower absorber-loop, forced-oxidation cases.
- Separate absorber and venturi hold tanks per train for venturi - spray tower processes with forced oxidation.

The oxidation areas for processes with forced oxidation differ mainly in tank configuration depending on the location of the tank.

The description of the reheat area is the same for all cases except the 0.7% sulfur coal cases in which no reheat is required. The reheater is

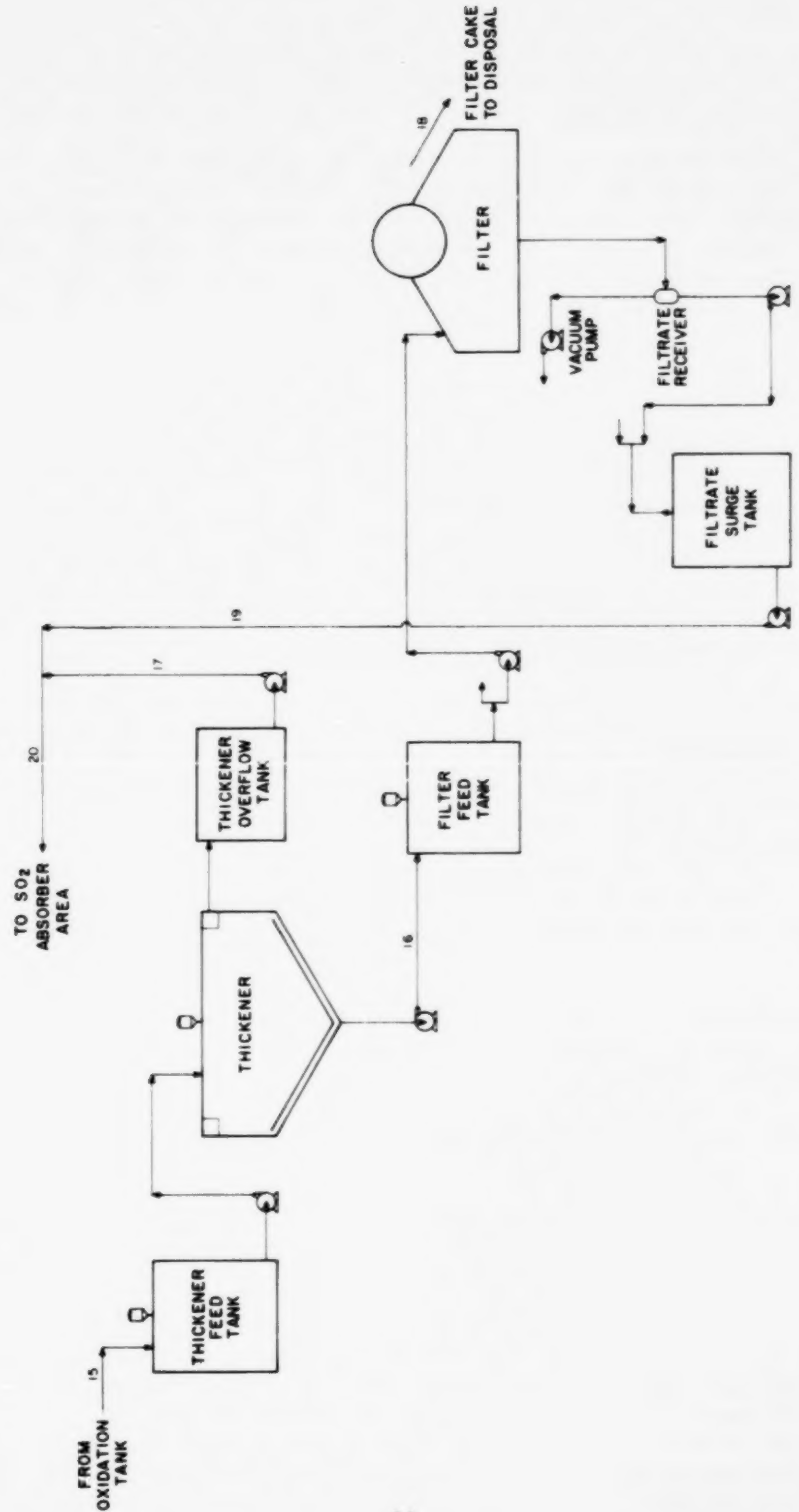

FROM OXIDATION TANK
15
THICKENER FEED TANK
THICKENER
16
FILTER FEED TANK
THICKENER OVERFLOW TANK
17
20
TO SO₂ ABSORBER AREA
19
FILTRATE SURGE TANK
FILTRATE RECEIVER
VACUUM PUMP
FILTER
18
FILTER CAKE TO DISPOSAL

Figure 5. Solids separation area, flow diagram.

essentially the same size for all cases at the same power unit size except that it is smaller for the adipic-acid-enhanced cases because partial bypass is used.

The solids separation and disposal area differ in general description only between ponding and landfill cases. The equipment sizes and quantities and the disposal site areas vary with waste quantity as affected by power unit size, coal sulfur content, and stoichiometry. In addition, processes with magnesium oxide additive have a 50% larger thickener area because of a slower settling rate, although the final filtered solids content is the same.

BASIC SYSTEMS

Because of the similarity of many of the case variations three limestone processes and one lime process are used to illustrate the various process options. The limestone processes are: case 6, TCA with ponding; case 15, spray tower with forced oxidation and landfill disposal; and case 22, venturi - spray tower with forced oxidation in the venturi hold tank and landfill disposal. The lime process is case 52, venturi - spray tower with magnesium oxide addition, bleedstream forced oxidation, and landfill disposal. Major operating conditions are shown in Table 16.

These processes cover the range of options examined in this study. The TCA with onsite ponding is similar to the limestone process evaluated in earlier FGD economic evaluations of this series (1,2). The spray tower with forced oxidation and landfill disposal illustrates current limestone technology. The third limestone process combines the venturi - spray tower absorber with forced oxidation in the venturi liquid loop. The lime process covers the use of lime absorbent and chemical additives along with forced oxidation of the bleedstream. Magnesium oxide was chosen instead of adipic acid because magnesium oxide is more effective in the bleedstream oxidation mode.

The equipment required for adipic acid addition is essentially the same as that for magnesium oxide addition; therefore, it also serves to illustrate the use of adipic acid.

Limestone Process With a TCA, Without Forced Oxidation, and Pond Disposal (Case 6)

The flow diagram and material balance for this process are shown in Figure 6 and Table 17 respectively. A plan and elevation drawing is shown in Figure 7 and an equipment list is shown in Table 18.

Limestone delivered to the site is unloaded by conveyor and stored in a 30-day stockpile. The limestone passes through a live storage hopper, which is built into the base of the storage pile, to a feed conveyor that transports it to the crushers. It is crushed and wet ground to 90% minus 325 mesh in two parallel ball mills and pumped as a 60% solids slurry to the slurry feed tank at the FGD unit. The fresh limestone slurry is metered into the absorber recirculation tank at a rate that maintains the solids concentration in the

TABLE 16. SUMMARY OF OPERATING CONDITIONS

Case	Process	L/G, gal/1,000 aft^3/min	Alkali stoichiometry, mols Ca/mol (SO_2 + 2HCl) absorbed	Gas velocity, ft/sec	Air stoichiometry, atom O/mol SO_2 absorbed	Residence time, min, hold tank/ oxidation tank	Sulfite oxidation, %	Filter cake solids, wt %	Pond settled solids, wt %
	Limestone Processes								
6	TCA - onsite ponding	58	1.4	12.5	2.5	8	30	-	40
15	Spray tower - forced oxidation - landfill disposal	106	1.4	10	2.5	8/5	95	85	-
22	Venturi - spray tower - forced oxidation - landfill disposal	20/103	1.3	10	2.5	8/5	95	85	-
	Lime Process								
52	Venturi - spray tower - MgO addition - bleed stream - forced oxidation - landfill disposal	20/40	1.01	10	2.5	8/5	95	85	-

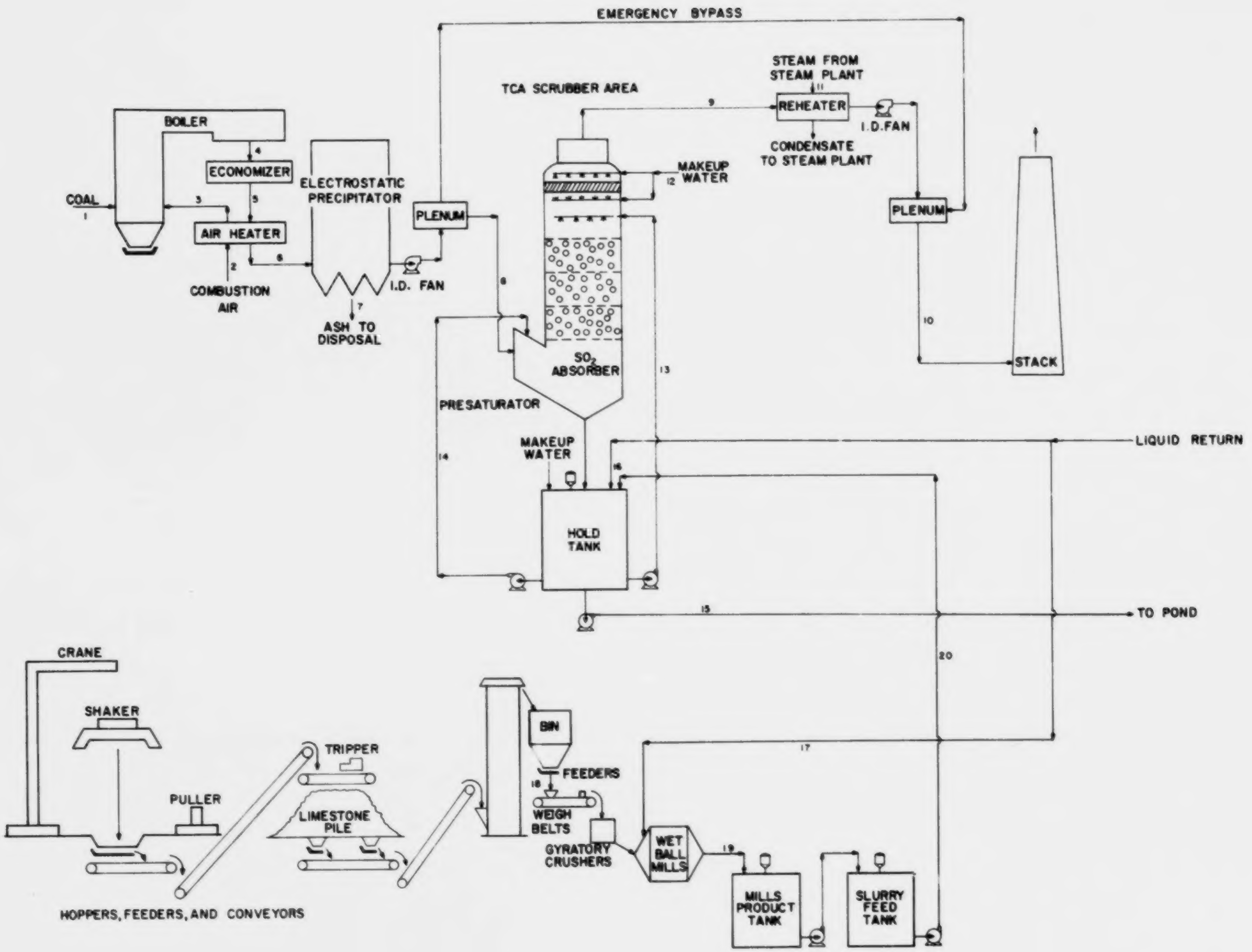

EMERGENCY BYPASS
STEAM FROM STEAM PLANT
TCA SCRUBBER AREA
REHEATER
I.D. FAN
CONDENSATE TO STEAM PLANT
BOILER
ECONOMIZER
COAL
AIR HEATER
COMBUSTION AIR
ELECTROSTATIC PRECIPITATOR
PLENUM
I.D. FAN
ASH TO DISPOSAL
MAKEUP WATER
PLENUM
STACK
SO_2 ABSORBER
PRESATURATOR
MAKEUP WATER
HOLD TANK
LIQUID RETURN
TO POND
CRANE
SHAKER
PULLER
TRIPPER
LIMESTONE PILE
HOPPERS, FEEDERS, AND CONVEYORS
BIN
FEEDERS
WEIGH BELTS
GYRATORY CRUSHERS
WET BALL MILLS
MILLS PRODUCT TANK
SLURRY FEED TANK
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
20

Figure 6. Limestone process with a TCA, without forced oxidation, and pond disposal, flow diagram.

TABLE 17. LIMESTONE PROCESS WITH A TCA, WITHOUT FORCED OXIDATION, AND POND DISPOSAL

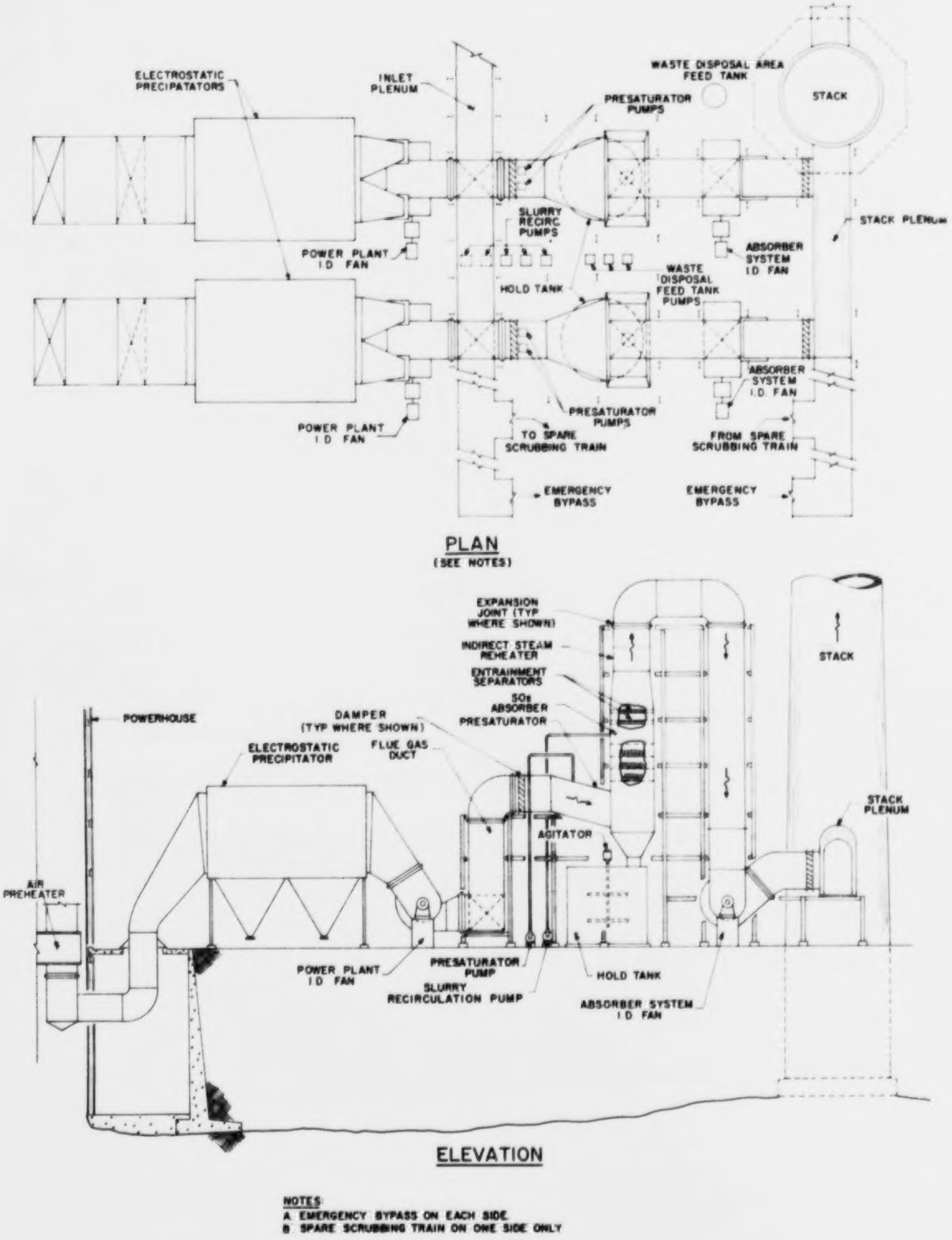
MATERIAL BALANCE

Stream No.		1	2	3	4	5
Description		Coal to boiler	Total air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,072,000	4,378,000	4,772,000	4,772,000
2						
3	Flow rate, sft^3/min@60°F		1,120,000	967,300	1,003,000	1,003,000
4	Temperature, $^\circ$F		80	495		705
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

Stream No.		6	7	8	9	10
Description		Gas to electrostatic precipitator	Fly ash to disposal	Gas to presaturator	Gas to reheater	Gas to stack
1	Total stream, lb/hr	5,465,000	50,050	5,415,000	5,637,000	5,643,000
2						
3	Flow rate, sft^3/min@60°F	1,157,000		1,156,000	1,235,000	1,237,000
4	Temperature, $^\circ$F	300		300	126	175
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

Stream No.		11	12	13	14	15
Description		Steam to reheater	Makeup water to absorber	Recycle slurry to absorber	Recycle slurry to presaturator	Absorber slurry to waste disposal
1	Total stream, lb/hr	92,200	408,800	42,310,000	2,918,000	838,300
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, $^\circ$F	470				
5	Pressure, psig	500				
6	Flow rate, gal/min		818	80,485	5,551	1,595
7	Specific gravity		1.0	1.05	1.05	1.05
8	pH					
9	Undissolved solids, %			8	8	8
10						

Stream No.		16	17	18	19	20
Description		Clear liquid return to scrubber	Clear liquid to wet ball mills	Limestone to weigh feeder	Slurry to mills product tank	Limestone slurry to effluent hold tank
1	Total stream, lb/hr	569,200	32,870	53,060	88,430	88,430
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, $^\circ$F					
5	Pressure, psig					
6	Flow rate, gal/min	1,139	66		112	112
7	Specific gravity	1.0	1.0		1.6	1.6
8	pH					
9	Undissolved solids, %				60	60
10						


ELECTROSTATIC PRECIPATATORS
INLET PLENUM
WASTE DISPOSAL AREA FEED TANK
STACK
PRESATURATOR PUMPS
SLURRY RECIRC PUMPS
STACK PLENUM
POWER PLANT I D FAN
ABSORBER SYSTEM I.D FAN
HOLD TANK
WASTE DISPOSAL FEED TANK PUMPS
ABSORBER SYSTEM I.D FAN
PRESATURATOR PUMPS
POWER PLANT I D FAN
TO SPARE SCRUBBING TRAIN
FROM SPARE SCRUBBING TRAIN
EMERGENCY BYPASS
EMERGENCY BYPASS
EXPANSION JOINT (TYP WHERE SHOWN)
INDIRECT STEAM REHEATER
ENTRAINMENT SEPARATORS
STACK
POWERHOUSE
DAMPER (TYP WHERE SHOWN)
SO_2 ABSORBER
PRESATURATOR
ELECTROSTATIC PRECIPITATOR
FLUE GAS DUCT
STACK PLENUM
AGITATOR
AIR PREHEATER
POWER PLANT I D FAN
PRESATURATOR PUMP
SLURRY RECIRCULATION PUMP
HOLD TANK
ABSORBER SYSTEM I D FAN

PLAN
(SEE NOTES)

ELEVATION

NOTES:
A. EMERGENCY BYPASS ON EACH SIDE.
B. SPARE SCRUBBING TRAIN ON ONE SIDE ONLY

Figure 7. Limestone process with a TCA, without forced oxidation, and pond disposal, plan and elevation.

recirculating slurry at 8%. Recycled pond water is also added to the recirculating tank to maintain the system water balance.

TABLE 18. LIMESTONE PROCESS WITH A TCA, WITHOUT FORCED OXIDATION, AND POND DISPOSAL

EQUIPMENT LIST

Item (number): description		Total material cost, 1982
Area 1--Materials Handling		
1.	Car shaker (1): Top mounting with crane, 20 hp shaker, 7-1/2 hp hoist	71,900
2.	Car puller (1): 25 hp puller, 5 hp return	63,100
3.	Hopper, limestone unloading (1): 16 ft diameter x 10 ft straight side, 2,570 ft^3, carbon steel, 30° cone bottom	15,500
4.	Feeder, limestone unloading (1): Vibrating pan, 30 in. wide x 60 in. long, 3-1/2 hp, 250 tons/hr	5,500
5.	Conveyor, limestone unloading (1): Belt, 36 in. wide x 20 ft long, 5 hp, 250 tons/hr, 165 ft/min	11,400
6.	Conveyor, limestone unloading (incline) (1): Belt, 36 in. wide x 310 ft long, 50 hp, 15° incline, 250 tons/hr, 165 ft/min	85,300
7.	Dust collector, limestone unloading pit (1): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning, includes dust hood	11,200
8.	Pump, limestone pit sump (1): Duplex, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	2,400
9.	Conveyor storage (1): Belt, 36 in. wide x 200 ft long, 5 hp, 250 tons/hr, 165 ft/min	73,100
10.	Tripper, storage conveyor (1): 1 hp, 30 ft/min	27,200
11.	Mobile equipment (1): Scraper tractor	141,900
12.	Hopper, reclaim (2): 7 ft diameter x 4-1/4 ft deep x 2 ft bottom, 75 ft^3, carbon steel, 60° cone bottom	2,400

(continued)

TABLE 18 (continued)

Item (number): description	Total material cost, 1982
13. Feeder, reclaim (2): Vibrating pan, 3-1/2 hp, 100 tons/hr	10,900
14. Dust collector, limestone reclaim pit (1): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning	7,800
15. Pump, reclaim pit sump (1): Duplex, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	2,400
16. Conveyor, limestone reclaim (1): Belt, 30 in. wide x 200 ft long, 5 hp, 100 tons/hr, 105 ft/min	40,900
17. Conveyor, limestone reclaim (incline) (1): Belt, 30 in. x 200 ft long, 40 hp, 15° incline, 45 ft lift, 100 tons/hr, 100 ft/min	60,300
18. Elevator, live limestone feed (1): Continuous bucket, 14 in. x 8 in. x 11-3/4 in., 75 hp, 90 ft lift, 100 tons/hr	57,800
19. Conveyor, live limestone feed (1): Belt, 30 in. wide x 60 ft long, 7-1/2 hp, 100 tons/hr, 100 ft/min	20,500
20. Tripper, live limestone feed conveyor (1): 1 hp, 30 ft/min	27,200
21. Bin, crusher feed (3): 13 ft diameter x 21 ft straight side height, 3,100 ft^3, w/cover, 50° slope, carbon steel (2 operating, 1 spare)	43,300
Subtotal	782,000
Area 2--Feed Preparation	
1. Feeder, crusher (3): Weigh belt, 14 ft long, 2 hp (2 operating, 1 spare)	49,600
2. Crusher (3): Gyratory, 75 hp (2 operating, 1 spare)	297,100
3. Ball mill (3): Wet, open system, 9-1/2 ft diameter x 19 ft long, 750 hp, 325 tons/day (2 operating, 1 spare)	1,733,800

(continued)

TABLE 18 (continued)

Item (number): description	Total material cost, 1982
4. *Dust collector, ball mill* (3): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning (2 operating, 1 spare)	23,300
5. *Tank, mills product* (3): 10 ft diameter x 10 ft high, 5,500 gal, open top, four 10 in. baffles, agitator supports, carbon steel, flakeglass lined (2 operating, 1 spare)	13,700
6. *Agitator, mills product tank* (3): 40 in. diameter, 10 hp, neoprene coated (2 operating, 1 spare)	22,900
7. *Pump, mills product tank* (3): Centrifugal, 55 gpm, 60 ft head, 2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	8,500
8. *Tank, slurry feed* (1): 22 ft diameter x 22 ft high, 58,900 gal, open top, four 22 in. baffles, agitator supports carbon steel, flakeglass lined	20,100
9. *Agitator, slurry feed* (1): 84 in. diameter, 49 hp, neoprene coated	43,600
10. *Pump, slurry feed* (8): Centrifugal, 27 gpm, 60 ft head, 1 hp, carbon steel, neoprene lined (4 operating, 4 spare)	21,500
Subtotal	2,234,100
Area 3--Gas Handling	
1. *Fans* (5): Induced draft, 377,300 aft^3/min, 15.2 in. static head, 1,275 hp, fluid drive, double width, double inlet, Inconel 625 (4 operating, 1 spare)	4,463,300
Subtotal	4,463,300
Area 4--SO_2 Absorption	
1. *SO_2 absorber* (5): Mobile bed absorber, 40 ft high x 34 ft wide x 17 ft deep, 1/4 in. carbon steel, neoprene lining, 316 stainless steel grids, nitrile foam spheres, FRP chevron vane entrainment separator, FRP spray headers (4 operating, 1 spare)	5,350,100

(continued)

TABLE 18 (continued)

Item (number): description	Total material cost, 1982
2. Tank, hold (5): 31 ft diameter x 31 ft high, 177,100 gal, open top, four 31 in. wide baffles, agitator supports, carbon steel, flakeglass lined (4 operating, 1 spare)	240,000
3. Agitator, hold tank (5): 93 in. diameter, 44 hp, neoprene coated (4 operating, 1 spare)	289,200
4. Pump, presaturator (10): Centrifugal, 1,400 gpm, 100 ft head, 61 hp, carbon steel, neoprene lined (4 operating, 6 spare)	103,200
5. Pump, recirculation (15): Centrifugal, 18,400 gpm, 100 ft head, 813 hp, carbon steel, neoprene lined (8 operating, 7 spare)	1,245,200
6. Pump, makeup water (2): Centrifugal, 2,775 gpm, 200 ft head, 234 hp, carbon steel (1 operating, 1 spare)	28,300
7. Soot blowers (40): Air, fixed	111,900
Subtotal	8,260,600
Area 5--Reheat	
1. Reheaters (5): Steam, tube type, 2,560 ft^2, six-tenths of tubes made of Inconel 625 and four-tenths made of Cor-Ten (4 operating, 1 spare)	2,557,000
2. Soot blowers (20): Air, retractable	183,000
Subtotal	2,740,000
Area 6--Oxidation	
Not applicable to this process	
Area 7--Solids Disposal	
1. Tank, pond feed (1): 16 ft diameter x 32 ft high, 48,100 gal, open top, four 16 in. baffles, agitator supports, carbon steel, flakeglass lined	19,100

(continued)

TABLE 18 (continued)

Item (number): description	Total material cost, 1982
2. Agitator, pond feed tank (1): 2 turbines, 48 in. diameter, 28 hp, neoprene coated	20,500
3. Pump, pond feed (2): Centrifugal, 1,600 gpm, 130 ft head, 90 hp, carbon steel, neoprene lined (1 operating, 1 spare)	23,800
4. Pump, pond return (2): Centrifugal, 1,200 gpm, 192 ft head 96 hp, carbon steel (1 operating, 1 spare)	17,000
Subtotal	80,400

A common feed plenum downstream of the power plant ESP units and ID fans distributes the flue gas to the five (one spare) absorber trains. Each train has a booster ID fan downstream of the reheaters. The flue gas enters the presaturator section of the TCA where it is cooled to 127°F by recycled scrubbing slurry. The flue gas then flows upward through the TCA countercurrent to the absorbent liquid.

The flue gas then passes through the mist eliminator section and enters the reheater section where the temperature is raised from 127°F to 169°F. The remaining reheat to 175°F is supplied by compression of the gas as it passes through the fans.

A bleedstream of spent scrubbing slurry overflows from the recirculation tank and flows by gravity to the pond feed tank. From this tank it is pumped one mile to the disposal pond. The waste slurry settles to a solids concentration of 40%, and the supernate is recycled to the wet ball mills and the recirculation tanks.

Limestone Process with a Spray Tower Absorber, Forced Oxidation, and Landfill Disposal (Case 15)

A flow diagram is shown in Figure 8 and the material balance is shown in Table 19. (A flow diagram of the solids separation area was previously shown in Figure 5.) A plan and elevation drawing is shown in Figure 9 and the equipment list is shown in Table 20.

Except for the use of a spray tower instead of a TCA and inclusion of forced oxidation and landfill instead of pond disposal, this process is similar to case 6 above. The materials handling, feed preparation, gas handling, and reheat areas are the same as those described for case 6. The SO_2 absorption area consists of a spray tower absorber. Air is introduced through a sparging ring into the oxidation tank beneath the spray tower at a rate to

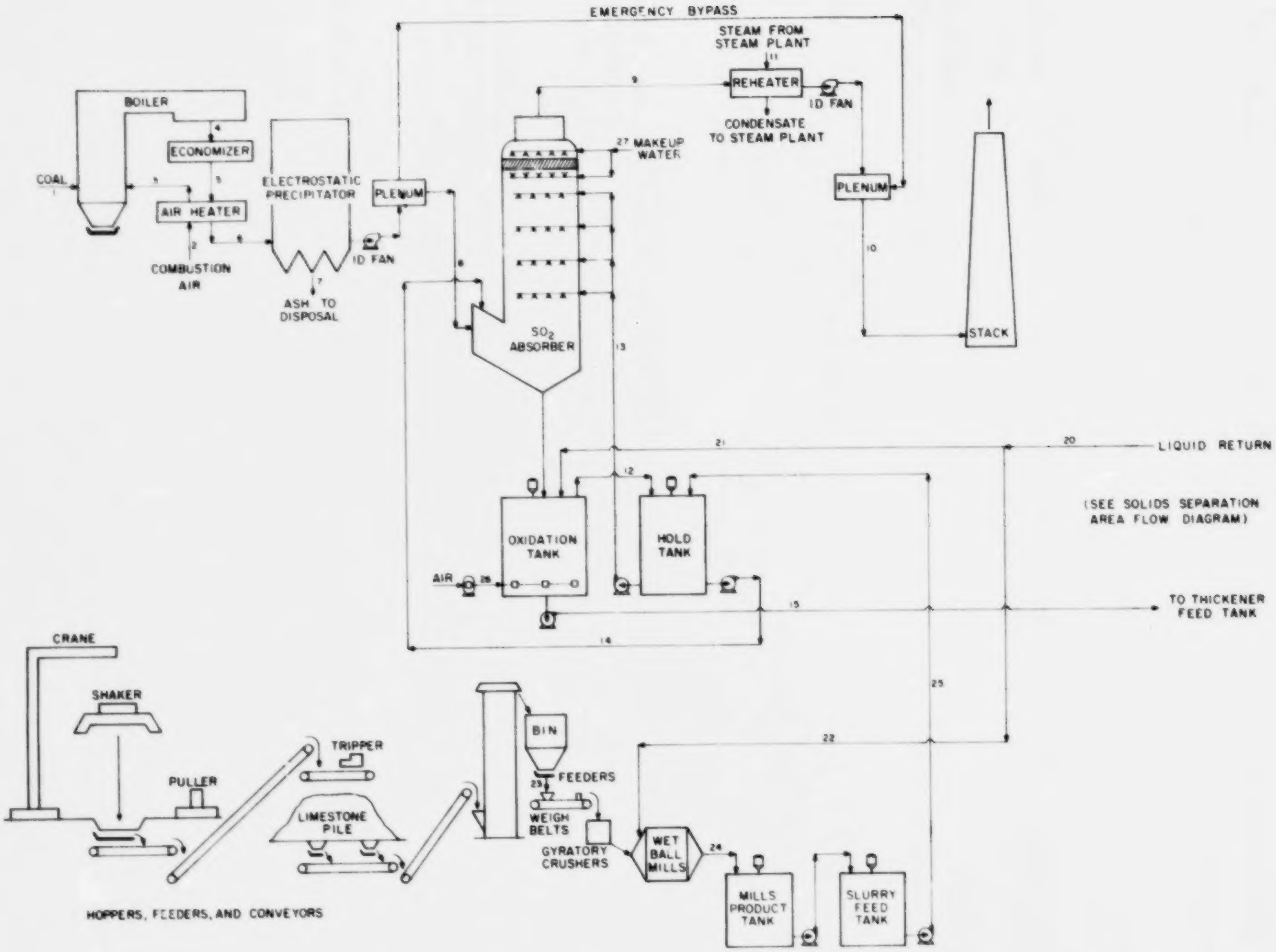

EMERGENCY BYPASS
STEAM FROM STEAM PLANT
11
REHEATER
9
ID FAN
CONDENSATE TO STEAM PLANT
BOILER
4
ECONOMIZER
27 MAKEUP WATER
COAL
1
3
5
ELECTROSTATIC PRECIPITATOR
PLENUM
PLENUM
AIR HEATER
2
6
ID FAN
COMBUSTION AIR
7
ASH TO DISPOSAL
8
10
SO$_2$ ABSORBER
13
STACK
21
20
LIQUID RETURN
12
(SEE SOLIDS SEPARATION AREA FLOW DIAGRAM)
OXIDATION TANK
HOLD TANK
AIR
26
15
TO THICKENER FEED TANK
14
CRANE
SHAKER
25
TRIPPER
BIN
22
PULLER
LIMESTONE PILE
23
FEEDERS
WEIGH BELTS
GYRATORY CRUSHERS
WET BALL MILLS
24
MILLS PRODUCT TANK
SLURRY FEED TANK
HOPPERS, FEEDERS, AND CONVEYORS

Figure 8. Limestone process with a spray tower absorber, forced oxidation, and landfill disposal, flow diagram.

TABLE 19. LIMESTONE PROCESS WITH A SPRAY TOWER ABSORBER, FORCED OXIDATION, AND LANDFILL DISPOSAL

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Total air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,072,000	4,378,000	4,772,000	4,772,000
2						
3	Flow rate, sft^3/min@60°F		1,120,000	967,300	1,003,000	1,003,000
4	Temperature, °F		80	495	705	705
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to electrostatic precipitator	Fly ash to disposal	Gas to presaturator	Gas to reheater	Gas to stack
1	Total stream, lb/hr	5,465,000	50,050	5,415,000	5,639,000	5,639,000
2						
3	Flow rate, sft^3/min@60°F	1,157,000		1,156,000	1,234,000	1,236,000
4	Temperature, °F	300		300	124	175
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	11	12	13	14	15
	Description	Steam to reheater	Oxidation tank overflow	Recycle slurry to absorber	Recycle slurry to presaturator	Absorber slurry to thickener
1	Total stream, lb/hr	98,580	80,100,000	77,270,000	2,916,000	948,900
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F	470				
5	Pressure, psig	500				
6	Flow rate, gal/min		152,600	147,000	5,547	1,805
7	Specific gravity		1.05	1.05	1.05	1.05
8	pH					
9	Undissolved solids, %		8	8	8	8
10						

	Stream No.	16	17	18	19	20
	Description	Thickener bottoms to filter	Thickener overflow return	Gypsum cake to disposal	Filtrate return	Clear liquid return
1	Total stream, lb/hr	189,800	731,600	89,300	100,900	832,100
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min	286	1,464		201	1,665
7	Specific gravity	1.3	1.0	2.0	1.0	1.0
8	pH					
9	Undissolved solids, %	40		85		
10						

(continued)

TABLE 19. (Continued)

	Stream No.	21	22	23	24	25
	Description	Clear liquid to oxidation tank	Clear liquid to wet ball mill	Limestone to weigh feeder	Slurry to mills product tank	Limestone slurry to recirculation tank
1	Total stream, lb/hr	800,000	32,130	51,950	86,580	86,580
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min	1,601	64		109	109
7	Specific gravity	1.0	1.0		1.6	1.6
8	pH					
9	Undissolved solids, %				60	60
10						

	Stream No.	26	27			
	Description	Air to oxidizer	Makeup water to absorber			
1	Total stream, lb/hr	59,950	286,900			
2						
3	Flow rate, sft^3/min@60°F	13,060				
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min		574			
7	Specific gravity		1.0			
8	pH					
9	Undissolved solids, %					
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

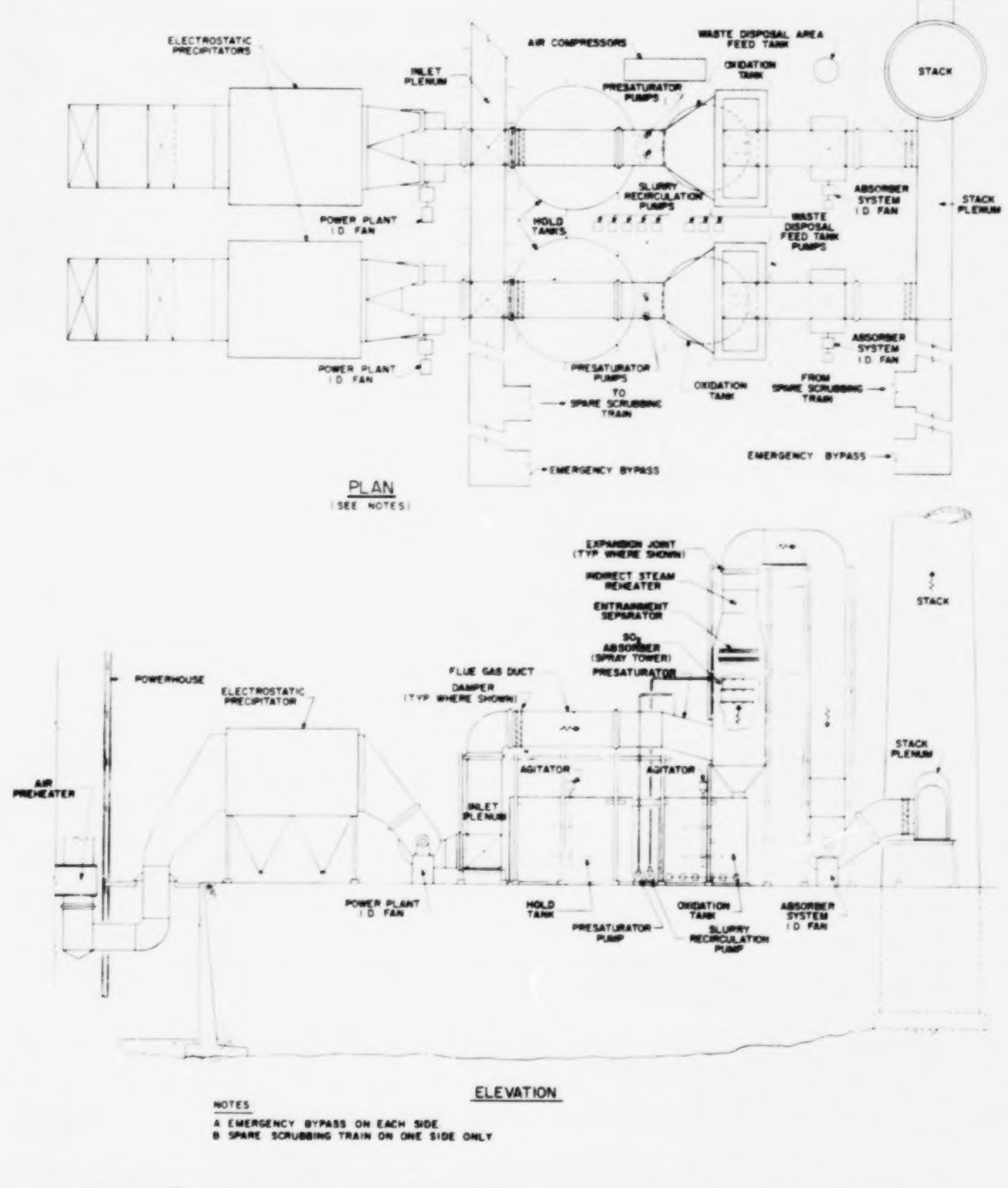

ELECTROSTATIC PRECIPITATORS
AIR COMPRESSORS
WASTE DISPOSAL AREA FEED TANK
INLET PLENUM
OXIDATION TANK
STACK
PRESATURATOR PUMPS
SLURRY RECIRCULATION PUMPS
ABSORBER SYSTEM I D FAN
STACK PLENUM
POWER PLANT I D FAN
HOLD TANKS
WASTE DISPOSAL FEED TANK PUMPS
ABSORBER SYSTEM I D FAN
POWER PLANT I D FAN
PRESATURATOR PUMPS
OXIDATION TANK
FROM SPARE SCRUBBING TRAIN
TO SPARE SCRUBBING TRAIN
EMERGENCY BYPASS
EMERGENCY BYPASS
PLAN
(SEE NOTES)
EXPANSION JOINT (TYP WHERE SHOWN)
INDIRECT STEAM REHEATER
ENTRAINMENT SEPARATOR
STACK
SO2 ABSORBER (SPRAY TOWER)
PRESATURATOR
FLUE GAS DUCT
POWERHOUSE
DAMPER (TYP WHERE SHOWN)
ELECTROSTATIC PRECIPITATOR
AGITATOR
AGITATOR
STACK PLENUM
AIR PREHEATER
INLET PLENUM
POWER PLANT I D FAN
HOLD TANK
PRESATURATOR PUMP
OXIDATION TANK
SLURRY RECIRCULATION PUMP
ABSORBER SYSTEM I D FAN
ELEVATION
NOTES
A EMERGENCY BYPASS ON EACH SIDE
B SPARE SCRUBBING TRAIN ON ONE SIDE ONLY

Figure 9. Limestone process with a spray tower absorber, forced oxidation, and landfill disposal, plan and elevation.

provide 2.5 atoms O/mol SO_2 absorbed. A separate hold tank is provided for makeup slurry addition. A bleedstream is withdrawn from the bottom of the oxidation tank and pumped to the thickener. The thickener underflow, at 35% to 40% solids, is dewatered to 85% solids in two rotary vacuum filters.

TABLE 20. LIMESTONE PROCESS WITH A SPRAY TOWER ABSORBER,

FORCED OXIDATION, AND LANDFILL DISPOSAL

EQUIPMENT LIST

Item (number): description	Total material cost, 1982
Area 1--Materials Handling	
1. Car shaker (1): Top mounting with crane, 20 hp shaker, 7-1/2 hp hoist	71,900
2. Car puller (1): 25 hp puller, 5 hp return	63,100
3. Hopper, limestone unloading (1): 16 ft diameter x 10 ft straight side, 2,570 ft^3, carbon steel, 30° cone bottom	15,500
4. Feeder, limestone unloading (1): Vibrating pan, 30 in. wide x 60 in. long, 3-1/2 hp, 250 tons/hr	5,500
5. Conveyor, limestone unloading (1): Belt, 36 in. wide x 20 ft long, 5 hp, 250 tons/hr, 165 ft/min	11,400
6. Conveyor, limestone unloading (incline) (1): Belt, 36 in. wide x 310 ft long, 50 hp, 15° incline, 250 tons/hr, 165 ft/min	85,300
7. Dust collector, limestone unloading pit (1): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning, includes dust hood	11,200
8. Pump, limestone pit sump (1): Duplex, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	2,400
9. Conveyor, storage (1): Belt, 36 in. wide x 200 ft long, 5 hp, 250 tons/hr, 165 ft/min	73,100
10. Tripper, storage conveyor (1): 1 hp, 30 ft/min	27,200
11. Mobile equipment (1): Scraper tractor	141,900

(continued)

TABLE 20 (continued)

Item (number): description	Total material cost, 1982
12. Hopper, reclaim (2): 7 ft diameter x 4-1/4 ft deep x 2 ft bottom, 75 ft^3, carbon steel, 60^o cone bottom	2,400
13. Feeder, reclaim (2): Vibrating pan, 3-1/2 hp, 100 tons/hr	10,900
14. Dust collector, limestone reclaim pit (1): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning	7,800
15. Pump, reclaim pit sump (1): Duplex, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	2,400
16. Conveyor, limestone reclaim (1): Belt, 30 in. wide x 200 ft long, 5 hp, 100 tons/hr, 105 ft/min	40,900
17. Conveyor, limestone reclaim (incline) (1): Belt, 30 in. wide x 200 ft long, 40 hp, 15^o incline, 45 ft lift, 100 tons/hr, 100 ft/min	60,300
18. Elevator, live limestone feed (1): Continuous bucket, 14 in. x 8 in. x 11-3/4 in., 75 hp, 90 ft lift, 100 tons/hr	57,800
19. Conveyor, live limestone feed (1): Belt, 30 in. wide x 60 ft long, 7-1/2 hp, 100 tons/hr, 100 ft/min	20,500
20. Tripper, live limestone feed conveyor (1): 1 hp, 30 ft/min	27,200
21. Bin, crusher feed (3): 13 ft diameter x 21 ft straight side height, 3,100 ft^3, w/cover, 50^o slope, carbon steel (2 operating, 1 spare)	43,300
Subtotal	782,000
Area 2--Feed Preparation	
1. Feeder, crusher (3): Weigh belt, 14 ft long, 2 hp (2 operating, 1 spare)	49,600
2. Crusher (3): Gyratory, 75 hp (2 operating, 1 spare)	297,100

(continued)

TABLE 20 (continued)

Item (number): description	Total material cost, 1982
3. Ball mill (3): Wet, open system, 9-1/2 ft diameter x 19 ft long, 740 hp, 325 tons/day (2 operating, 1 spare)	1,721,200
4. Dust collector, ball mill (3): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning (2 operating, 1 spare)	23,300
5. Tank, mills product (3): 10 ft diameter x 10 ft high, 5,500 gal, open top, four 10 in. baffles, agitator supports, carbon steel, flakeglass lined (2 operating, 1 spare)	13,700
6. Agitator, mills product tank (3): 40 in. diameter, 10 hp, neoprene coated (2 operating, 1 spare)	22,900
7. Pump, mills product tank (3): Centrifugal, 55 gpm, 60 ft head, 2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	8,500
8. Tank, slurry feed (1): 21 ft diameter x 21 ft high, 57,700 gal, open top, four 21 in. baffles, agitator supports, carbon steel, flakeglass lined	19,800
9. Agitator, slurry feed (1): 84 in. diameter, 49 hp, neoprene coated	43,600
10. Pump, slurry feed (8): Centrifugal, 27 gpm, 60 ft head, 1 hp, carbon steel, neoprene lined (4 operating, 4 spare)	21,500
Subtotal	2,221,200
Area 3--Gas Handling	
1. Fans (5): Induced draft, 377,300 aft^3/min, 7.9 in. static head, 663 hp, fluid drive, double width, double inlet, Inconel 625 (4 operating, 1 spare)	3,401,300
Subtotal	3,401,300

(continued)

TABLE 20 (continued)

Item (number): description	Total material cost, 1982
Area 4--SO_2 Absorption	
1. SO_2 absorber (5): Spray tower, 40 ft high x 34 ft wide x 17 ft deep, 1/4 in. carbon steel, neoprene lining, 316 stainless steel grids, FRP chevron vane entrainment separator (4 operating, 1 spare)	5,517,200
2. Tank, hold (5): 38 ft diameter x 38 ft high, 323,400 gal, open top, four 38 in. wide baffles, agitator supports, carbon steel, flakeglass lined (4 operating, 1 spare)	358,500
3. Agitator, hold tank (5): 152 in. diameter, 66 hp, neoprene coated (4 operating, 1 spare)	442,800
4. Pump, presaturator (10): Centrifugal, 1,400 gpm, 100 ft head, 61 hp, carbon steel, neoprene lined (4 operating, 6 spare)	103,200
5. Pump, recirculation (15): Centrifugal, 18,400 gpm, 100 ft head, 813 hp, carbon steel, neoprene lined (8 operating, 7 spare)	1,689,800
6. Pump, makeup water (2): Centrifugal, 3,470 gpm, 200 ft head, 292 hp, carbon steel (1 operating, 1 spare)	33,200
7. Soot blowers (40): Air, fixed	111,900
Subtotal	8,256,600
Area 5--Reheat	
1. Reheaters (5): Steam, tube type, 2,750 ft^2, six-tenths of tubes made of Inconel 625 and four-tenths made of Cor-Ten (4 operating, 1 spare)	2,638,800
2. Soot blowers (20): Air, retractable	183,000
Subtotal	2,821,800

(continued)

TABLE 20 (continued)

Item (number): description	Total material cost, 1982
Area 6--Oxidation	
1. Tank, oxidation (5): 30 ft diameter x 38 ft high, 202,100 gal, open top, four 30 in. wide baffles, agitator supports, carbon steel, flakeglass lined, includes air sparger (4 operating, 1 spare)	366,700
2. Agitator, oxidation tank (5): 120 in. diameter, 59 hp, neoprene coated (4 operating, 1 spare)	317,600
3. Pump, oxidation tank (8): Centrifugal, 450 gpm, 60 ft head, 12 hp, carbon steel, neoprene lined (4 operating, 4 spare)	37,700
4. Blower, oxidation air (6): 3,264 sft^3/min, 322 hp (4 operating, 2 spare)	208,400
Subtotal	930,400
Area 7--Solids Separation and Disposal	
1. Tank, thickener feed (1): 19 ft diameter x 38 ft high, 85,760 gal, open top, four 19 in. wide baffles, agitator supports, carbon steel, flakeglass lined	28,500
2. Agitator, thickener feed tank (1): 80 in. diameter, 41 hp, neoprene coated	31,300
3. Pump, thickener feed (2): Centrifugal, 1,800 gpm, 60 ft head, 48 hp, carbon steel, neoprene lined (1 operating, 1 spare)	19,200
4. Thickener (1): 48 ft diameter x 5-1/2 ft high, carbon steel sides, concrete basin, includes 1 hp rake motor and mechanism	85,500
5. Tank, thickener overflow (1): 27-1/2 ft diameter x 5-1/2 ft high, 24,150 gal, open top, carbon steel	6,600

(continued)

TABLE 20 (continued)

Item (number): description	Total material cost, 1982
6. Pump, thickener overflow tank (2): Centrifugal, 1,464 gpm, 75 ft head, 46 hp, carbon steel, neoprene lined (1 operating, 1 spare)	12,900
7. Pump, thickener underflow (2): Centrifugal, 286 gpm, 9 ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	7,800
8. Tank, filter feed (1): 9 ft diameter x 9 ft high, 4,725 gal, open top, four 9 in. wide baffles, agitator supports, carbon steel, flakeglass lined	3,700
9. Agitator, filter feed tank (1): 36 in. diameter, 7 hp, neoprene coated	5,600
10. Pump, filter feed tank (3): Centrifugal, 143 gpm, 50 ft head, 4 hp, carbon steel, neoprene lined (2 operating, 1 spare)	11,900
11. Filter (3): Rotary vacuum, 8 ft diameter x 14 ft face, 85 total hp, includes auxiliary equipment (2 operating, 1 spare)	380,800
12. Pump, filtrate (4): Centrifugal, 100 gpm, 20 ft head, 1 hp, carbon steel, neoprene lined (2 operating, 2 spare)	17,300
13. Tank, filtrate surge (1): 8 ft diameter x 8 ft high, 3,320 gal, open top, carbon steel	1,700
14. Pump, filtrate surge tank (2): Centrifugal, 200 gpm, 85 ft head, 7 hp, carbon steel, neoprene lined (1 operating, 1 spare)	9,300
15. Conveyor, filter cake (incline) (1): Belt, 30 in. wide x 175 ft long, 1-1/2 hp, 100 ft incline, 50 tons/hr	37,100
16. Retaining wall, gypsum (1): 8 ft high x 32 ft long, concrete	1,600
Subtotal	660,800

The filter cake is transferred outside the filter building by conveyor and placed in a temporary storage area on the ground. It is then loaded on trucks by a front-end loader and transported one mile to a permanent landfill. A description of the landfill is given in the premises.

Limestone Process with a Venturi - Spray Tower Absorber, Forced Oxidation, and Landfill Disposal (Case 22)

A flow diagram is shown in Figure 10 and the material balance is shown in Table 21. (A flow diagram of the solids separation area was previously shown in Figure 5.) A plan and elevation is shown in Figure 11. The equipment list is shown in Table 22. The venturi serves as a presaturator and removes 20% of the SO_2 in the flue gas. The flue gas passes from the venturi to the spray tower, which is similar to the previously described spray tower except for a lower L/G ratio.

Makeup slurry is added to the spray tower hold tank. The spray tower slurry overflows to the venturi hold tank, which operates at a lower pH (about 4.8). The venturi hold tank is used as the oxidation tank. Compressed air is introduced into it through a sparging ring at a rate to provide 2.5 atoms O/mol SO_2 absorbed. A bleedstream containing about 8% solids is pumped from the tank to the dewatering area where it is thickened and filtered to 85% solids and disposed of in a landfill.

Lime Process with a Venturi - Spray Tower Absorber, MgO Addition, Bleedstream Forced Oxidation, and Landfill Disposal (Case 52)

This process illustrates the use of lime instead of limestone absorbent, the use of a chemical additive to enhance SO_2 removal, and forced oxidation in the bleedstream rather than within the scrubbing loop. The flow diagram is shown in Figure 12 and the material balance is shown in Table 23. (A flow diagram of the solids separation area was previously shown in Figure 5.) A plan and elevation diagram is shown in Figure 13 and the equipment list is shown in Table 24.

Magnesium oxide addition appears particularly attractive for the bleedstream oxidation process because it enhances SO_2 removal in the scrubbing loop and also buffers the slurry against a pH rise in the oxidation tank caused by residual limestone. Within-loop forced oxidation in a magnesium-oxide-enhanced system has not been proven effective (27). One disadvantage of magnesium oxide addition is that the oxidized slurry settling rate is adversely affected. Results at Shawnee indicate that the settling rate is reduced to as low as 50% of the settling rate exhibited by an oxidized slurry without magnesium oxide addition (27), a factor compensated for in the dewatering area design.

The raw material handling and preparation areas for a lime process are somewhat different from a limestone process. The lime is received by truck or rail and an enclosed conveyor transports the lime to concrete storage silos with a 30-day capacity. The lime is then conveyed as needed to a feed bin from which it is metered into a dual-compartment overflow slaker. A 20% solids slurry is produced and pumped to the slurry feed tank. A 30-day supply of magnesium oxide is maintained in a storage silo adjacent to the slaker

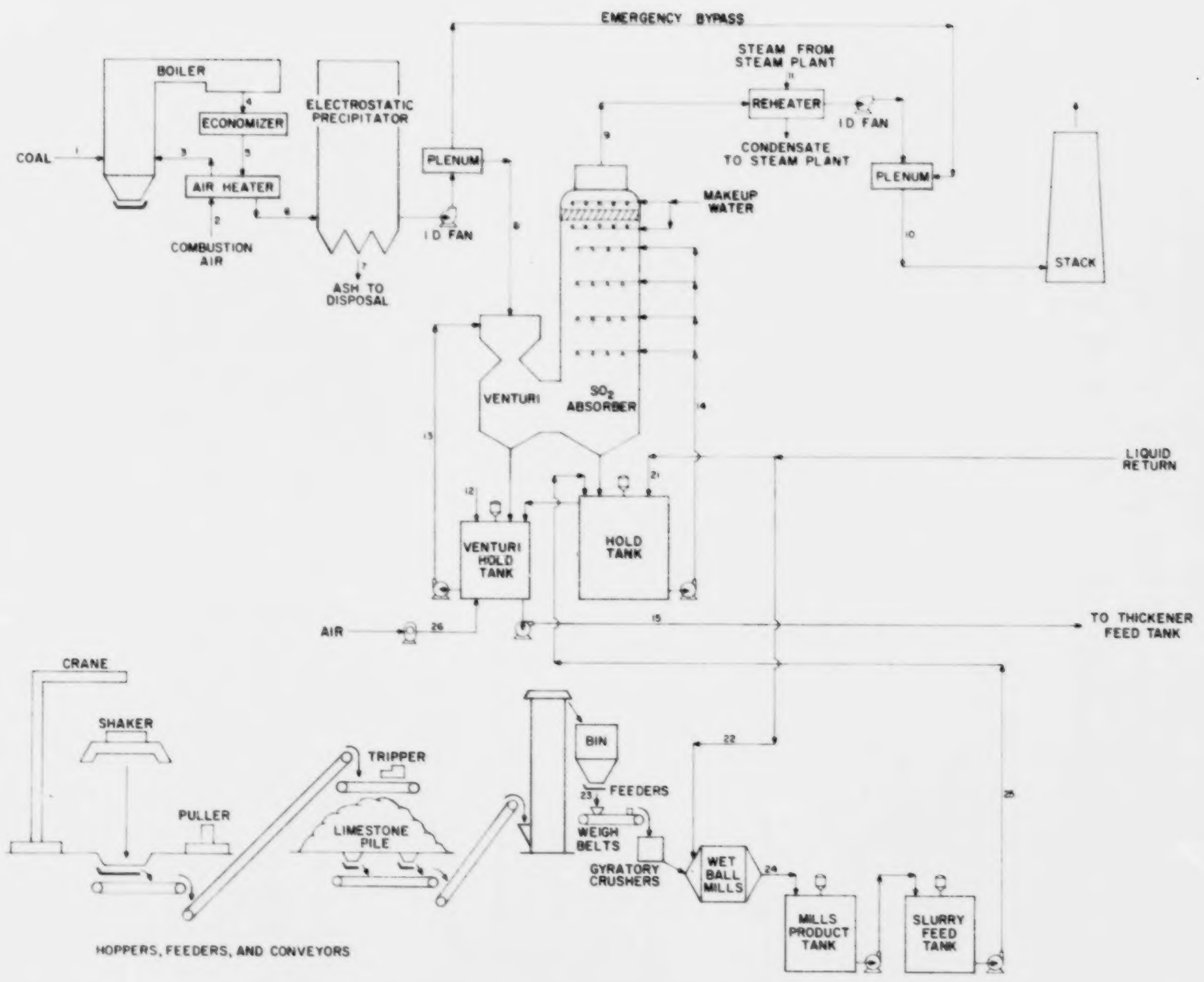

EMERGENCY BYPASS
STEAM FROM STEAM PLANT
BOILER
REHEATER
ELECTROSTATIC PRECIPITATOR
ECONOMIZER
I D FAN
COAL
PLENUM
CONDENSATE TO STEAM PLANT
AIR HEATER
PLENUM
MAKEUP WATER
COMBUSTION AIR
I D FAN
STACK
ASH TO DISPOSAL
VENTURI
SO2 ABSORBER
LIQUID RETURN
VENTURI HOLD TANK
HOLD TANK
AIR
TO THICKENER FEED TANK
CRANE
SHAKER
TRIPPER
BIN
FEEDERS
PULLER
LIMESTONE PILE
WEIGH BELTS
GYRATORY CRUSHERS
WET BALL MILLS
MILLS PRODUCT TANK
SLURRY FEED TANK
HOPPERS, FEEDERS, AND CONVEYORS
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
21
22
23
24
25
26

Figure 10. Limestone process with a venturi - spray tower absorber, forced oxidation, and landfill disposal, flow diagram.

TABLE 21. LIMESTONE PROCESS WITH A VENTURI - SPRAY TOWER ABSORBER, FORCED OXIDATION, AND LANDFILL DISPOSAL

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Total air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,077,000	4,378,000	4,772,000	4,772,000
2						
3	Flow rate, sft^3/min@60°F		1,120,000	967,300	1,003,000	1,003,000
4	Temperature, °F		80	605	705	705
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to electrostatic precipitator	Fly ash to disposal	Gas to venturi - spray tower	Gas to reheater	Gas to stack
1	Total stream, lb/hr	5,665,000	50,050	5,615,000	5,639,000	5,639,000
2						
3	Flow rate, sft^3/min@60°F	1,157,000		1,156,000	1,234,000	1,236,000
4	Temperature, °F	300		300	126	175
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	11	12	13	14	15
	Description	Steam to reheater	Makeup water	Recycle slurry to venturi	Recycle slurry to tower	Oxidized slurry to thickener
1	Total stream, lb/hr	89,550	284,900	14,580,000	75,080,000	902,400
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F	470				
5	Pressure, psig	500				
6	Flow rate, gal/min		570	27,740	142,800	1,717
7	Specific gravity			1.05	1.05	1.05
8	pH					
9	Undissolved solids, %			8	8	8
10						

	Stream No.	16	17	18	19	20
	Description	Thickener bottoms to filter	Thickener overflow return	Gypsum cake to disposal	Filtrate return	Clear liquid return
1	Total stream, lb/hr	180,500	695,800	84,930	95,550	791,400
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min	272	1,392		191	1,583
7	Specific gravity	1.3	1.0	2.0	1.0	1.0
8	pH					
9	Undissolved solids, %	40		85		
10						

(continued)

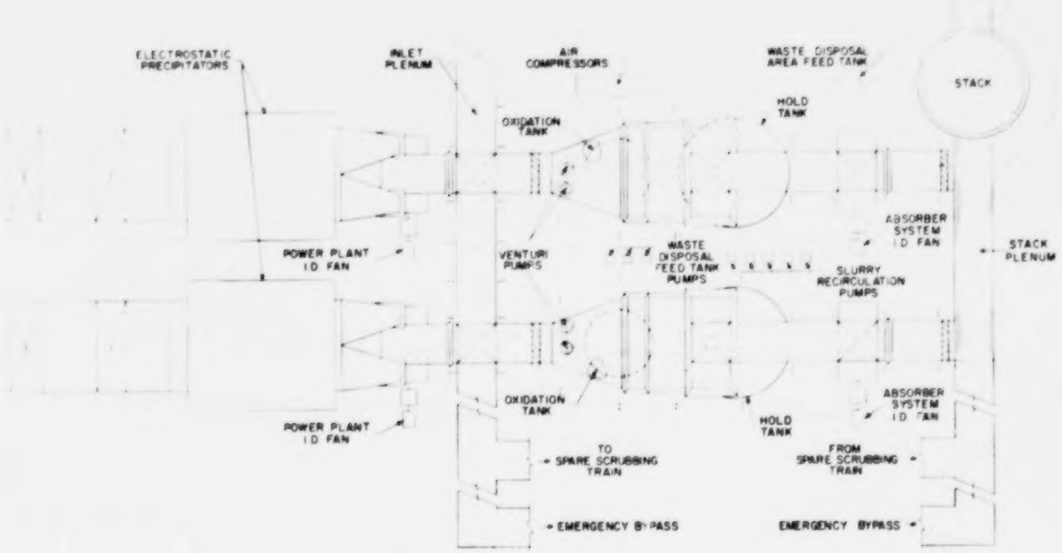
TABLE 21. (Continued)

	Stream No.	21	22	23	24	25
	Description	Clear liquid return to scrubber	Clear liquid return to ball mill	Limestone to weigh feeder	Slurry to mills product tank	Limestone slurry to recirculation tank
1	Total stream, lb/hr	761,200	30,160	48,240	80,390	80,390
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psia					
6	Flow rate, gal/min	1,523	60		101	101
7	Specific gravity	1.0	1.0		1.6	1.6
8	pH					
9	Undissolved solids, %				60	60
10						

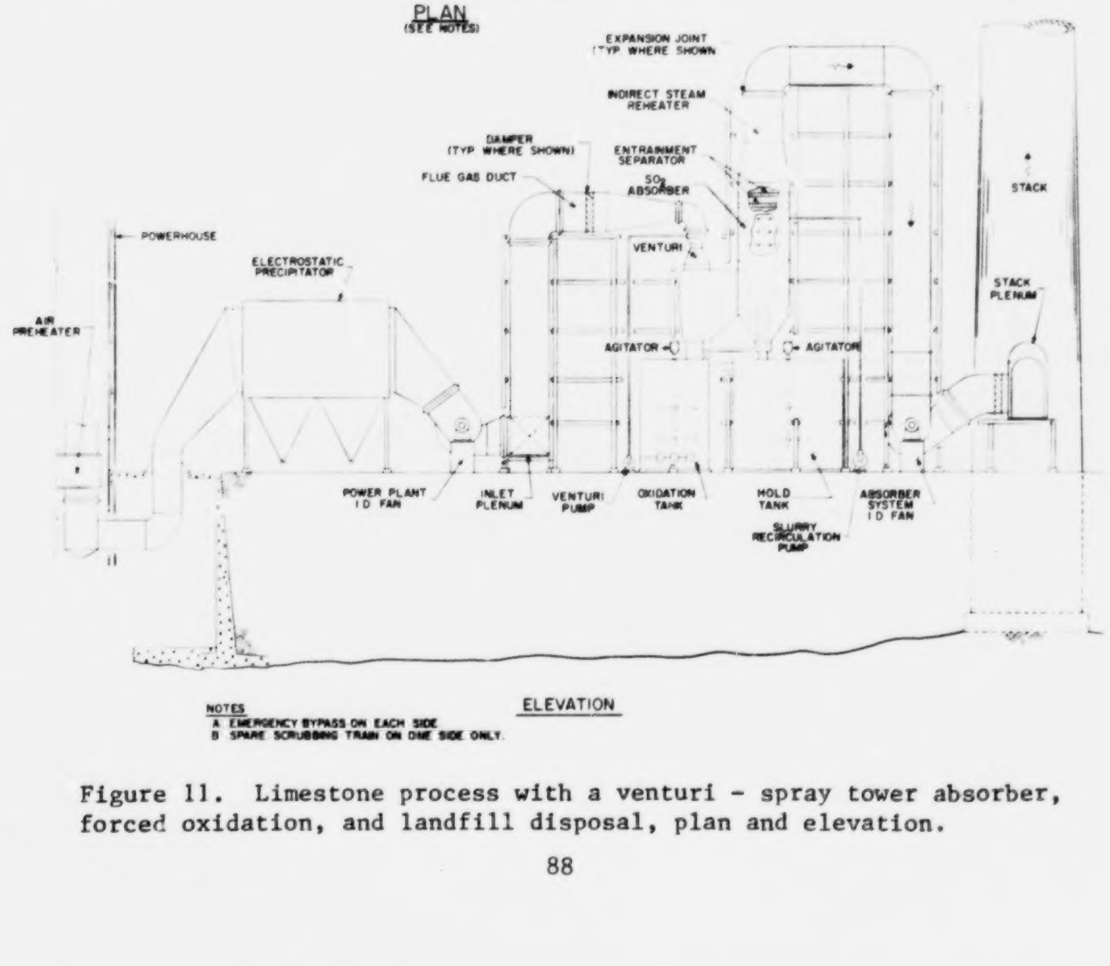
	Stream No.	26				
	Description	Air to oxidiser				
1	Total stream, lb/hr	59,950				
2						
3	Flow rate, sft^3/min@60°F	13,060				
4	Temperature, °F					
5	Pressure, psia					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						


ELECTROSTATIC PRECIPITATORS
INLET PLENUM
AIR COMPRESSORS
WASTE DISPOSAL AREA FEED TANK
STACK
OXIDATION TANK
HOLD TANK
ABSORBER SYSTEM I.D FAN
STACK PLENUM
POWER PLANT I.D FAN
VENTURI PUMPS
WASTE DISPOSAL FEED TANK PUMPS
SLURRY RECIRCULATION PUMPS
OXIDATION TANK
HOLD TANK
ABSORBER SYSTEM I.D FAN
POWER PLANT I.D FAN
TO SPARE SCRUBBING TRAIN
FROM SPARE SCRUBBING TRAIN
EMERGENCY BYPASS
EMERGENCY BYPASS

PLAN
(SEE NOTES)


EXPANSION JOINT (TYP WHERE SHOWN)
INDIRECT STEAM REHEATER
DAMPER (TYP WHERE SHOWN)
ENTRAINMENT SEPARATOR
FLUE GAS DUCT
SO2 ABSORBER
STACK
POWERHOUSE
VENTURI
ELECTROSTATIC PRECIPITATOR
STACK PLENUM
AIR PREHEATER
AGITATOR
AGITATOR
POWER PLANT I D FAN
INLET PLENUM
VENTURI PUMP
OXIDATION TANK
HOLD TANK
ABSORBER SYSTEM I D FAN
SLURRY RECIRCULATION PUMP

ELEVATION

NOTES
A EMERGENCY BYPASS ON EACH SIDE
B SPARE SCRUBBING TRAIN ON ONE SIDE ONLY.

Figure 11. Limestone process with a venturi - spray tower absorber, forced oxidation, and landfill disposal, plan and elevation.

area. It is unloaded and transferred pneumatically. The magnesium oxide is metered into the slaker product tanks from a feed bin located next to the slakers to maintain a total Mg^{2+} concentration of 5,500 ppm in the absorbent.

TABLE 22. LIMESTONE PROCESS WITH A VENTURI - SPRAY TOWER ABSORBER, FORCED OXIDATION, AND LANDFILL DISPOSAL

EQUIPMENT LIST

Item (number): description	Total material cost, 1982
Area 1--Materials Handling	
1. Car shaker (1): Top mounting with crane, 20 hp shaker, 7-1/2 hp hoist	71,900
2. Car puller (1): 25 hp puller, 5 hp return	63,100
3. Hopper, limestone unloading (1): 16 ft diameter x 10 ft straight side, 2,570 ft^3, carbon steel, 30° cone bottom	15,500
4. Feeder, limestone unloading (1): Vibrating pan, 30 in. wide x 60 in. long, 3-1/2 hp, 250 tons/hr	5,500
5. Conveyor, limestone unloading (1): Belt, 36 in. wide x 20 ft long, 5 hp, 250 tons/hr, 165 ft/min	11,400
6. Conveyor, limestone unloading (incline) (1): Belt, 36 in. wide x 310 ft long, 50 hp, 15° incline, 250 tons/hr, 165 ft/min	85,300
7. Dust collector, limestone unloading pit (1): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning, includes dust hood	11,200
8. Pump, limestone pit sump (1): Duplex, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	2,400
9. Conveyor, storage (1): Belt, 36 in. wide x 200 ft long, 5 hp, 250 tons/hr, 165 ft/min	73,100
10. Tripper, storage conveyor (1): 1 hp, 30 ft/min	27,200
11. Mobile equipment (1): Scraper tractor	141,900

(continued)

TABLE 22 (continued)

Item (number): description	Total material cost, 1982
12. Hopper, reclaim (2): 7 ft diameter x 4-1/4 ft deep x 2 ft bottom, 75 ft^3, carbon steel, 60° cone bottom	2,400
13. Feeder, reclaim (2): Vibrating pan, 3-1/2 hp, 100 tons/hr	10,900
14. Dust collector, limestone reclaim pit (1): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning	7,800
15. Pump, reclaim pit sump (1): Duplex, 60 gpm, 70 ft head, 5 hp, carbon steel, neoprene lined	2,400
16. Conveyor, limestone reclaim (1): Belt, 30 in. wide x 200 ft long, 5 hp, 100 tons/hr, 100 ft/min	40,900
17. Conveyor, limestone reclaim (incline) (1): Belt, 30 in. wide x 193 ft long, 40 hp, 15° incline, 45 ft lift, 100 tons/hr, 100 ft/min	60,300
18. Elevator, live limestone feed (1): Continuous bucket, 14 in. x 8 in. x 11-3/4 in., 75 hp, 90 ft lift, 100 tons/hr	57,800
19. Conveyor, live limestone feed (1): Belt, 30 in. wide x 60 ft long, 7-1/2 hp, 100 tons/hr, 100 ft/min	20,500
20. Tripper, live limestone feed conveyor (1): 1 hp, 30 ft/min	27,200
21. Bin, crusher feed (3): 13 ft diameter x 21 ft straight side height, 3,100 ft^3, w/cover, 50° slope, carbon steel (2 operating, 1 spare)	43,300
Subtotal	782,000
Area 2--Feed Preparation	
1. Feeder, crusher (3): Weigh belt, 14 ft long, 2 hp (2 operating, 1 spare)	49,600
2. Crusher (3): Gyratory, 75 hp (2 operating, 1 spare)	297,100

(continued)

TABLE 22 (continued)

Item (number): description	Total material cost, 1982
3. Ball mill (3): Wet, open system, 9 ft diameter x 19 ft long, 687 hp, 300 tons/day (2 operating, 1 spare)	1,677,800
4. Dust collector, ball mill (3): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning (2 operating, 1 spare)	23,300
5. Tank, mills product (3): 10 ft diameter x 10 ft high, 5,500 gal, open top, four 10 in. baffles, agitator supports, carbon steel, flakeglass lined (2 operating, 1 spare)	13,700
6. Agitator, mills product tank (3): 40 in. diameter, 10 hp, neoprene coated (2 operating, 1 spare)	22,900
7. Pump, mills product tank (3): Centrifugal, 51 gpm, 60 ft head, 2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	8,500
8. Tank, slurry feed (1): 21 ft diameter x 21 ft high, 53,600 gal, open top, four 21 in. baffles, agitator supports, carbon steel, flakeglass lined	18,900
9. Agitator, slurry feed (1): 82 in. diameter, 47 hp, neoprene coated	40,700
10. Pump, slurry feed (8): Centrifugal, 25 gpm, 60 ft head, 1 hp, carbon steel, neoprene lined (4 operating, 4 spare)	21,500
Subtotal	2,174,000
Area 3--Gas Handling	
1. Fans (5): Induced draft, 377,300 aft^3/min, 18.0 in. static head, 1,511 hp, fluid drive, double width, double inlet, Inconel 625 (4 operating, 1 spare)	4,855,600
Subtotal	4,855,600

(continued)

TABLE 22 (continued)

Item (number): description	Total material cost, 1982
Area 4--SO_2 Absorption	
1. SO_2 absorber (5): Venturi scrubber with acid-brick-lined throat (4 operating, 1 spare)	792,800
SO_2 absorber (5): Spray tower, 40 ft high x 34 ft wide x 17 ft deep, 1/4 in. carbon steel, neoprene lining, 316 stainless steel grids, FRP chevron vane entrainment separator (4 operating, 1 spare)	5,501,100
2. Tank, hold (5): 38 ft diameter x 38 ft high, 314,200 gal, open top, four 38 in. wide baffles, agitator supports, carbon steel, flakeglass lined (4 operating, 1 spare)	351,700
3. Agitator, hold tank (5): 152 in. diameter, 65 hp, neoprene coated (4 operating, 1 spare)	433,900
4. Pump, venturi recirculation (10): Centrifugal, 6,950 gpm, 100 ft head, 307 hp, carbon steel, neoprene lined (4 operating, 6 spare)	788,000
5. Pump, spray tower recirculation (15): Centrifugal, 17,850 gpm, 100 ft head, 790 hp, carbon steel, neoprene lined (8 operating, 7 spare)	1,662,100
6. Pump, makeup water (2): Centrifugal, 3,470 gpm, 200 ft head, 292 hp, carbon steel (1 operating, 1 spare)	33,200
7. Soot blowers (55): Air, fixed	153,900
Subtotal	9,716,700
Area 5--Reheat	
1. Reheaters (5): Steam, tube type, 2,480 ft^2, six-tenths of tubes made of Inconel 625 and four-tenths made of Cor-Ten (4 operating, 1 spare)	2,522,400
2. Soot blowers (20): Air, retractable	183,000
Subtotal	2,705,400

(continued)

TABLE 22 (continued)

Item (number): description	Total material cost, 1982
Area 6--Oxidation	
1. Tank, oxidation (5): 19 ft diameter x 38 ft high, 78,470 gal, open top, four 19 in. wide baffles, agitator supports, carbon steel, flakeglass lined, includes air sparger (4 operating, 1 spare)	195,500
2. Agitator, oxidation tank (5): 76 in. diameter, 46 hp, neoprene coated (4 operating, 1 spare)	162,600
3. Pump, oxidation tank (8): Centrifugal, 429 gpm, 60 ft head, 11 hp, carbon steel, neoprene lined (4 operating, 4 spare)	37,400
4. Blower, oxidation air (6): 3,264 sft^3/min, 320 hp (4 operating, 2 spare)	208,400
Subtotal	603,900
Area 7--Solids Separation and Disposal	
1. Tank, thickener feed (1): 19 ft diameter x 38 ft high, 78,470 gal, open top, four 19 in. wide baffles, agitator supports, carbon steel, flakeglass lined	28,000
2. Agitator, thickener feed tank (1): 76 in. diameter, 41 hp, neoprene coated	30,700
3. Pump, thickener feed (2): Centrifugal, 1,720 gpm, 60 ft head, 46 hp, carbon steel, neoprene lined (1 operating, 1 spare)	19,000
4. Thickener (1): 46 ft diameter x 5-1/2 ft high, carbon steel sides, concrete basin, includes 1 hp rake motor and mechanism	83,000
5. Tank, thickener overflow (1): 27 ft diameter x 5-1/2 ft high, 22,970 gal, open top, carbon steel	6,300

(continued)

TABLE 22 (continued)

Item (number): description	Total material cost, 1982
6. Pump, thickener overflow tank (2): Centrifugal, 1,400 gpm, 75 ft head, 44 hp, carbon steel, neoprene lined (1 operating, 1 spare)	12,700
7. Pump, thickener underflow (2): Centrifugal, 272 gpm, 9 ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	7,800
8. Tank, filter feed (1): 9 ft diameter x 9 ft high, 4,500 gal, open top, four 9 in. wide baffles, agitator supports, carbon steel, flakeglass lined	3,600
9. Agitator, filter feed tank (1): 36 in. diameter, 7 hp, neoprene coated	5,300
10. Pump, filter feed tank (3): Centrifugal, 136 gpm, 50 ft head, 4 hp, carbon steel, neoprene lined (2 operating, 1 spare)	11,900
11. Filter (3): Rotary vacuum, 8 ft diameter x 14 ft face, 85 total hp, includes auxiliary equipment (2 operating, 1 spare)	372,300
12. Pump, filtrate (4): Centrifugal, 96 gpm, 20 ft head, 1 hp, carbon steel, neoprene lined (2 operating, 2 spare)	17,200
13. Tank, filtrate surge (1): 8 ft diameter x 8 ft high, 3,150 gal, open top, carbon steel	1,700
14. Pump, filtrate surge tank (2): Centrifugal, 191 gpm, 85 ft head, 6 hp, carbon steel, neoprene lined (1 operating, 1 spare)	9,200
15. Conveyor, filter cake (incline) (1): Belt, 30 in. wide x 175 ft long, 1-1/2 hp, 100 ft incline, 50 tons/hr	37,100
16. Retaining wall, gypsum (1): 8 ft high x 32 ft long, concrete	1,600
Subtotal	647,400

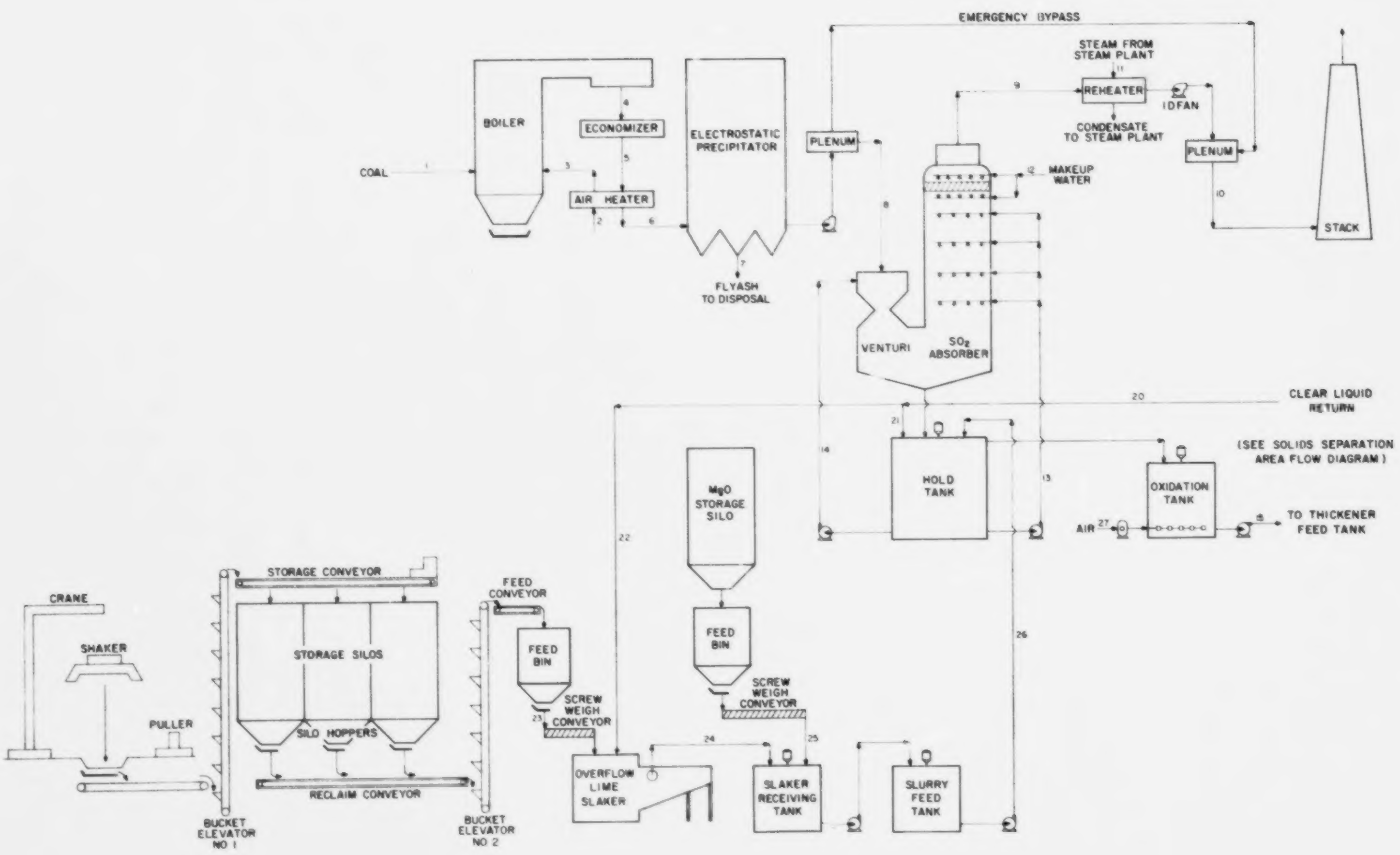

EMERGENCY BYPASS
STEAM FROM STEAM PLANT
11
9
REHEATER
IDFAN
CONDENSATE TO STEAM PLANT
PLENUM
PLENUM
12 MAKEUP WATER
10
STACK
BOILER
ECONOMIZER
4
5
COAL
1
3
AIR HEATER
2
6
ELECTROSTATIC PRECIPITATOR
7
FLYASH TO DISPOSAL
8
VENTURI
SO_2 ABSORBER
20
CLEAR LIQUID RETURN
21
14
HOLD TANK
13
(SEE SOLIDS SEPARATION AREA FLOW DIAGRAM)
OXIDATION TANK
AIR 27
15 TO THICKENER FEED TANK
22
MgO STORAGE SILO
FEED BIN
SCREW WEIGH CONVEYOR
26
CRANE
SHAKER
PULLER
STORAGE CONVEYOR
STORAGE SILOS
SILO HOPPERS
RECLAIM CONVEYOR
BUCKET ELEVATOR NO 1
FEED CONVEYOR
FEED BIN
SCREW WEIGH CONVEYOR
23
BUCKET ELEVATOR NO 2
OVERFLOW LIME SLAKER
24
25
SLAKER RECEIVING TANK
SLURRY FEED TANK

Figure 12. Lime process with a venturi - spray tower absorber, MgO addition, bleedstream forced oxidation, and landfill disposal, flow diagram.

TABLE 23. LIME PROCESS WITH A VENTURI - SPRAY TOWER ABSORBER, MgO ADDITION, BLEEDSTREAM FORCED OXIDATION, AND LANDFILL DISPOSAL

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Total air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,068,000	4,375,000	4,771,000	4,771,000
2						
3	Flow rate, sft^3/min@60°F		1,120,000	967,300	1,003,000	1,003,000
4	Temperature, °F		80	495	705	705
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	6	7	8	9	10
	Description	Gas to electrostatic precipitator	Fly ash to disposal	Gas to venturi - spray tower	Gas to reheater	Gas to stack
1	Total stream, lb/hr	5,465,000	50,050	5,415,000	5,639,000	5,639,000
2						
3	Flow rate, sft^3/min@60°F	1,157,000		1,136,000	1,234,000	1,236,000
4	Temperature, °F	300		300	124	125
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

	Stream No.	11	12	13	14	15
	Description	Steam to reheater	Makeup water	Recycle slurry to tower	Recycle slurry to venturi	Oxidized slurry to thickener
1	Total stream, lb/hr	90,690	178,400	29,090,000	14,550,000	758,500
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F	470				
5	Pressure, psig	500				
6	Flow rate, gal/min		357	55,345	27,672	1,443
7	Specific gravity		1.0	1.05	1.05	1.05
8	pH					
9	Undissolved solids, %			8	8	8
10						

	Stream No.	16	17	18	19	20
	Description	Thickener bottoms to filter	Thickener overflow return	Gypsum cake to disposal	Filtrate return	Clear liquid return
1	Total stream, lb/hr	151,700	584,900	71,390	80,320	665,300
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min	229	1,170		161	1,331
7	Specific gravity	1.3	1.0	2.0	1.0	1.0
8	pH					
9	Undissolved solids, %	40		85		
10						

(continued)

TABLE 23. (Continued)

	Stream No.	21	22	23	24	25
	Description	Clear liquid return to scrubber	Clear liquid return to slaker	Lime feed to weigh feeder	Slurry to slaker product tank	MgO to slaker product tank
1	Total stream, lb/hr	548,800	116,500	21,000	138,700	19
2						
3	Flow rate, sft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min	1,098	233		246	
7	Specific gravity	1.0	1.0		1.12	
8	pH					
9	Undissolved solids, %				20	
10						

	Stream No.	26	27			
	Description	Lime slurry to recirculation tank	Air to oxidizer			
1	Total stream, lb/hr	138,700	59,950			
2						
3	Flow rate, sft^3/min@60°F		13,060			
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min	246				
7	Specific gravity	1.12				
8	pH					
9	Undissolved solids, %	20				
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

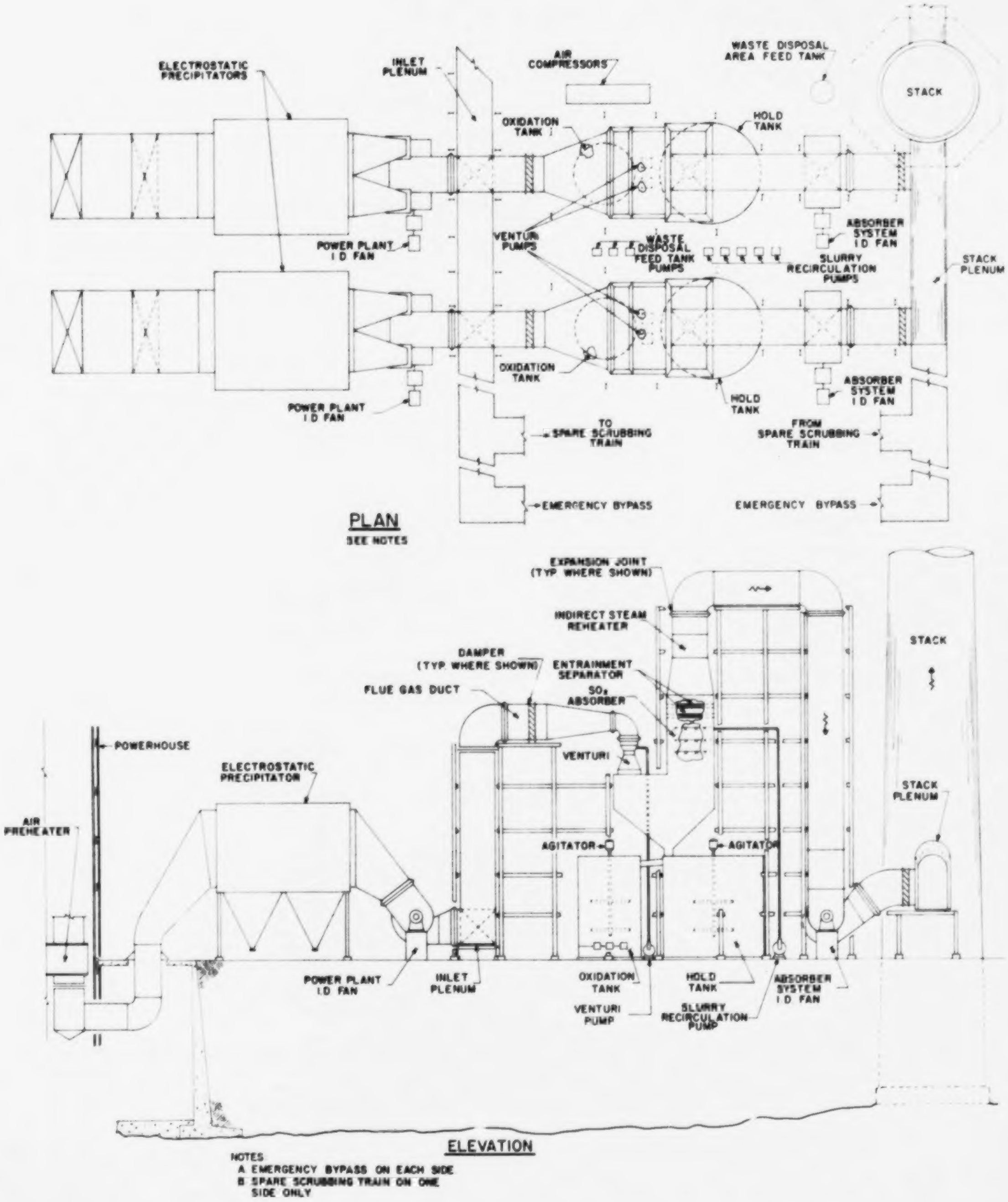

ELECTROSTATIC PRECIPITATORS
INLET PLENUM
AIR COMPRESSORS
WASTE DISPOSAL AREA FEED TANK
STACK
OXIDATION TANK
HOLD TANK
POWER PLANT I D FAN
VENTURI PUMPS
WASTE DISPOSAL FEED TANK PUMPS
SLURRY RECIRCULATION PUMPS
ABSORBER SYSTEM I D FAN
STACK PLENUM
OXIDATION TANK
HOLD TANK
ABSORBER SYSTEM I D FAN
POWER PLANT I D FAN
TO SPARE SCRUBBING TRAIN
FROM SPARE SCRUBBING TRAIN
EMERGENCY BYPASS
EMERGENCY BYPASS
PLAN
SEE NOTES
EXPANSION JOINT (TYP WHERE SHOWN)
INDIRECT STEAM REHEATER
STACK
DAMPER (TYP WHERE SHOWN)
ENTRAINMENT SEPARATOR
FLUE GAS DUCT
SO₂ ABSORBER
POWERHOUSE
VENTURI
ELECTROSTATIC PRECIPITATOR
AIR PREHEATER
STACK PLENUM
AGITATOR
AGITATOR
POWER PLANT I.D. FAN
INLET PLENUM
OXIDATION TANK
VENTURI PUMP
HOLD TANK
SLURRY RECIRCULATION PUMP
ABSORBER SYSTEM I.D. FAN
ELEVATION
NOTES
A EMERGENCY BYPASS ON EACH SIDE
B SPARE SCRUBBING TRAIN ON ONE SIDE ONLY

Figure 13. Lime process with a venturi - spray tower absorber, MgO addition, bleedstream forced oxidation, and landfill disposal, plan and elevation.

TABLE 24. LIME PROCESS WITH A VENTURI - SPRAY TOWER ABSORBER, MgO ADDITION, BLEEDSTREAM FORCED OXIDATION, AND LANDFILL DISPOSAL

EQUIPMENT LIST

Item (number): description	Total material cost, 1982
Area 1--Materials Handling	
1. Car shaker (1): Top mounting with crane, 20 hp shaker, 7-1/2 hp hoist	71,900
2. Car puller (1): 25 hp puller, 5 hp return	63,100
3. Hopper, lime unloading (1): 16 ft diameter x 10 ft straight side, 2,570 ft^3, carbon steel, 30° cone bottom	15,500
4. Feeder, lime unloading (1): Vibrating pan, 30 in. wide x 60 in. long, 3-1/2 hp, 100 tons/hr	3,800
5. Conveyor, lime unloading (1): Belt, 24 in. wide x 20 ft long, enclosed, 5 hp, 100 tons/hr	11,400
6. Dust collector, lime unloading pit (1): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning, includes dust hood	11,200
7. Elevator, storage silo (1): Continuous bucket, 16 in. x 13 in. x 18 in., 50 hp, 122 ft lift	75,500
8. Silo, lime storage (2): 49 ft diameter x 73 ft straight side height, 137,400 ft^3, concrete	341,700
9. Hopper, storage silo base (2): Carbon steel, 60° cone bottom	44,700
10. Feeder, reclaim (2): Vibrating pan, 3-1/2 hp	7,600
11. Conveyor, reclaim (1): Belt, 18 in. wide x 173 ft long, enclosed horizontally, 5 hp	32,000
12. Elevator, feed bin (1): Continuous bucket, 16 in. x 9 in. x 12 in., 50 hp, 50 ft lift	37,800

(continued)

TABLE 24 (continued)

Item (number): description	Total material cost, 1982
13. Conveyor, feed bin (1): Belt, 18 in. wide x 60 ft long, enclosed horizontally, 5 hp	15,900
14. Tripper, feed conveyor (1): 1 hp, 30 ft/min	18,900
15. Bin, lime feed (3): 10 ft diameter x 15 ft straight side height, 60° slope (2 operating, 1 spare)	14,500
Subtotal	765,500
Area 2--Feed Preparation	
1. Feeder, vibrating bin (3): 3.5 hp (2 operating, 1 spare)	30,500
2. Feeder, slaker (3): Screw conveyor, 12 in. diameter x 12 ft long, 1 hp (2 operating, 1 spare)	15,600
3. Slaker (3): Dual compartment, overflow style, 10 hp, 5.0 tons/hr (2 operating, 1 spare)	170,400
4. Dust collector, lime system (3): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning (2 operating, 1 spare)	7,800
5. Tank, slaker product (3): 12 ft diameter x 12 ft high, 5,500 gal, open top, four 12 in. baffles, agitator supports, carbon steel, flakeglass lined (2 operating, 1 spare)	15,400
6. Agitator, slaker product tank (3): 40 in. diameter, 10 hp, neoprene coated (2 operating, 1 spare)	22,900
7. Pump, slaker product tank (3): Centrifugal, 123 gpm, 60 ft head, 4 hp, carbon steel, neoprene lined (2 operating, 1 spare)	11,800
8. Silo, MgO storage (1): 6 ft diameter x 14 ft straight side height, 265 ft^3, carbon steel 60° cone bottom	4,000
9. Conveyor, MgO feed bin (1): Pneumatic, 10 hp	8,600

(continued)

TABLE 24 (continued)

Item (number): description	Total material cost, 1982
10. Bin, MgO feed (1): Carbon steel, rubber lined	1,300
11. Feeder, MgO addition (1): Screw conveyor, 6 in. diameter x 30 ft long, 316 stainless steel	4,700
12. Dust collector, additive (1): Bag filter, polypropylene bag, 450 ft^3/min, 1.5 hp	3,300
13. Tank, slurry feed (1): 28 ft diameter x 28 ft high, 130,000 gal, open top, four 28 in. baffles, agitator supports, carbon steel, flakeglass lined	34,100
14. Agitator, slurry feed (1): 112 in. diameter, 48 hp neoprene coated	53,900
15. Pump, slurry feed (8): Centrifugal, 62 gpm, 60 ft head, 2 hp, carbon steel, neoprene lined (4 operating, 4 spare)	22,400
Subtotal	406,700
Area 3--Gas Handling	
1. Fans (5): Induced draft, 377,300 aft^3/min, 16.5 in. static head, 1,382 hp, fluid drive, double width, double inlet, Inconel 625 (4 operating, 1 spare)	4,641,400
Subtotal	4,641,400
Area 4--SO$_2$ Absorption	
1. First-stage SO$_2$ absorber (5): Venturi with acid-brick-lined throat (4 operating, 1 spare)	792,800
2. Second-stage SO$_2$ absorber (5): Spray tower, 40 ft high x 34 ft wide x 17 ft deep, 1/4 in. carbon steel, neoprene lining, 316 stainless steel grids, FRP chevron vane entrainment separator, FRP spray headers (4 operating, 1 spare)	5,091,600

(continued)

TABLE 24 (continued)

Item (number): description	Total material cost, 1982
3. Tank, hold (5): 27-1/2 ft diameter x 27-1/2 ft high, 121,800 gal, open top, four 28 in. wide baffles, agitator supports, carbon steel, flakeglass lined (4 operating, 1 spare)	163,100
4. Agitator, hold tank (5): 110 in. diameter, 35 hp, neoprene coated (4 operating, 1 spare)	221,900
5. Pump, venturi recirculation (10): Centrifugal, 6,920 gpm, 100 ft head, 306 hp, carbon steel, neoprene lined (4 operating, 6 spare)	787,000
6. Pump, spray tower, recirculation (15): Centrifugal, 6,920 gpm, 100 ft head, 306 hp, carbon steel, neoprene lined (8 operating, 7 spare)	1,180,500
7. Pump, makeup water (2): Centrifugal, 3,460 gpm, 200 ft head, 291 hp, carbon steel (1 operating, 1 spare)	33,100
8. Soot blowers (55): Air, fixed	218,600
Subtotal	8,488,600
Area 5--Reheat	
1. Reheaters (5): Steam, tube type, 2,520 ft^2, six-tenths of tubes made of Inconel 625 and four-tenths made of Cor-Ten (4 operating, 1 spare)	2,536,000
2. Soot blowers (20): Air, retractable	183,000
Subtotal	2,719,000
Area 6--Oxidation	
1. Tank, oxidation (5): 14 ft diameter x 27-1/2 ft high, 30,400 gal, open top, four 14 in. wide baffles, agitator supports, carbon steel, flakeglass lined, includes air sparger (4 operating, 1 spare)	74,500

(continued)

TABLE 24 (continued)

Item (number): description	Total material cost, 1982
2. Agitator, oxidation tank (5): 56 in. diameter, 22 hp, neoprene coated (4 operating, 1 spare)	78,500
3. Pump, oxidation tank (8): Centrifugal, 361 gpm, 60 ft head, 10 hp, carbon steel, neoprene lined (4 operating, 4 spare)	36,400
4. Blower, oxidation air (6): 3,264 sft^3/min, 233 hp (4 operating, 2 spare)	208,400
Subtotal	397,800
Area 7--Solids Separation and Disposal	
1. Tank, thickener feed (1): 14 ft diameter x 28 ft high, 30,400 gal, open top, four 14 in. wide baffles, agitator supports, carbon steel, flakeglass lined	14,900
2. Agitator, thickener feed tank (1): 56 in. diameter, 22 hp, neoprene coated	15,700
3. Pump, thickener feed (2): Centrifugal, 1,443 gpm, 60 ft head, 38 hp, carbon steel, neoprene lined (1 operating, 1 spare)	18,200
4. Thickener (1): 60 ft diameter x 6 ft high, carbon carbon steel sides, concrete basin, includes 1 hp rake motor and mechanism	112,800
5. Tank, thickener overflow (1): 24 ft diameter x 6 ft high, 19,300 gal, open top, carbon steel	5,500
6. Pump, thickener overflow tank (2): Centrifugal, 1,170 gpm, 75 ft head, 37 hp, carbon steel, neoprene lined (1 operating, 1 spare)	12,100
7. Pump, thickener underflow (2): Centrifugal, 229 gpm, 9 ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	7,700

(continued)

TABLE 24 (continued)

Item (number): description	Total material cost, 1982
8. Tank, filter feed (1): 8-1/2 ft diameter x 8-1/2 ft high, 3,800 gal, open top, four 9 in. wide baffles, agitator supports, carbon steel, flakeglass lined	3,200
9. Agitator, filter feed tank (1): 36 in. diameter, 6 hp, neoprene coated	4,500
10. Pump, filter feed tank (3): Centrifugal, 114 gpm, 50 ft head, 3 hp, carbon steel, neoprene lined (2 operating, 1 spare)	11,700
11. Filter (3): Rotary vacuum, 8 ft diameter x 12 ft face, 38 total hp, includes auxiliary equipment (2 operating, 1 spare)	344,200
12. Pump, filtrate (4): Centrifugal, 80 gpm, 20 ft head, 1 hp, carbon steel, neoprene lined (2 operating, 2 spare)	17,100
13. Tank, filtrate surge (1): 8 ft diameter x 8 ft high, 2,500 gal, open top, carbon steel	1,400
14. Pump, filtrate surge tank (2): Centrifugal, 161 gpm, 85 ft head, 6 hp, carbon steel, neoprene lined (1 operating, 1 spare)	9,100
15. Conveyor, filter cake (incline) (1): Belt, 30 in. wide x 175 ft long, 1-1/2 hp, 100 ft incline, 50 tons/hr	37,100
16. Retaining wall, gypsum (1): 8 ft high x 32 ft long, concrete	1,600
Subtotal	616,800

In the bleedstream oxidation cases the venturi and spray tower drain into a single hold tank, where fresh lime slurry and clarified water are added. The L/G ratio in the spray tower is reduced to 40 gal/kaft^3 and the stoichiometry is 1.01. Design of the oxidation tank is similar to that for the cases previously described.

An overflow line from the hold tank transfers some of the slurry to an adjacent tank where it is oxidized by sparging air into the tank. The slurry is then pumped to the dewatering area for disposal, in the same manner as the other forced-oxidation systems.

RESULTS

The overall economic results consist of the capital investment in 1982 dollars, first-year annual revenue requirements (which include levelized capital charges) in 1984 dollars, and levelized annual revenue requirements (which consist of levelized operating and maintenance costs as well as levelized capital charges) also in 1984 dollars. The levelizing factor adjusts for 6% annual inflation and 10% annual cost of money over the 30-year life of the installation. The derivation of the economic results is discussed in more detail in the premise section.

COST SUMMARIES AND DETAILED COMPARISONS

Detailed costs for each case are shown in Appendixes A, B, C, and D. Summaries of these costs are shown in Tables 25, 26, 27, and 28. Comparisons are made between the three main classes of limestone processes--those without forced oxidation and those with forced oxidation, both with and without additives, and between lime and limestone processes. Except for two cases all forced-oxidation processes include landfill. All processes without forced oxidation (or natural oxidation) include pond disposal. Comparison of other factors such as absorber type, coal sulfur content, and power unit size is also made. The relationship of the annual revenue requirements of processes may vary because of variations in the cost of common components that compose different fractions of the totals or because of cost components unique to one process. The sensitivities of annual revenue requirements to these variations are also compared.

Comparison of Limestone Processes With and Without Forced Oxidation

Cost comparisons by process area for limestone processes without forced oxidation and limestone processes with forced oxidation at different coal sulfur contents are shown in Tables 29 and 30. The capital investment and levelized annual revenue requirements for the 500-MW cases at the various coal sulfur levels are shown in Figure 14. In all cases capital investments for the processes without forced oxidation are higher. The differences increase with increasing coal sulfur content (ranging from 4% higher for the low sulfur cases up to 13% higher for the 5.0% sulfur coal case). The higher capital investment is largely a result of disposal area construction costs, which are two to three-and-one-half times higher for ponds than for landfills. In contrast, the additional capital investment for forced oxidation and waste dewatering (solids separation) is relatively small.

TABLE 25. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS

FOR LIMESTONE PROCESSES WITHOUT FORCED OXIDATION

							Annual revenue requirements, 1984$			
		Plant size,	Coal sulfur,	Absorber	Capital investment, 1982$		First-year		Levelized	
Case	Absorbent	MW	%	type	M$	$/kW	M$	Mills/kWh	M$	Mills/kWh
1	Limestone	200	3.5	ST	62.7	314	15.8	14.39	21.7	19.72
2	Limestone	500	0.7	ST	73.2	146	18.1	6.56	24.5	8.91
3	Limestone	500	2.0	ST	90.4	181	23.3	8.46	32.1	11.68
4	Limestone	500	3.5	ST	112.6	225	28.8	10.47	39.6	14.42
5	Limestone	500	5.0	ST	123.3	247	31.5	11.44	43.3	15.73
6	Limestone	500	3.5	TCA	107.8	216	27.6	10.03	38.0	13.81
7	Limestone	500	3.5	VST	121.7	243	31.3	11.37	43.1	15.68
8	Limestone	1,000	3.5	ST	192.1	192	48.9	8.89	67.2	12.23

Note: No additives, pond disposal in all cases.

TABLE 26. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR LIMESTONE PROCESSES WITH FORCED OXIDATION

					Capital investment, 1982$		Annual revenue requirements, 1984$			
							First-year		Levelized	
Case	Absorbent	Plant size, MW	Coal sulfur, %	Absorber type	M$	$/kW	M$	Mills/kWh	M$	Mills/kWh
9	Limestone	200	0.7	ST	44.4	222	12.3	11.15	17.3	15.76
10	Limestone	200	2.0	ST	52.1	260	14.7	13.40	21.0	19.10
11	Limestone	200	3.5	ST	59.1	295	16.7	15.21	23.9	21.69
12	Limestone	200	5.0	ST	62.6	313	17.8	16.17	25.4	23.08
13	Limestone	500	0.7	ST	70.4	141	18.8	6.84	26.3	9.56
14	Limestone	500	2.0	ST	86.6	173	24.2	8.80	34.4	12.49
15	Limestone	500	3.5	ST	103.1	206	29.1	10.58	41.5	15.08
16	Limestone	500	5.0	ST	109.2	218	31.2	11.36	44.7	16.25
17	Limestone	500	3.5	TCA	98.5	197	27.8	10.09	39.5	14.37
18	Limestone	500	3.5	VST	109.2	218	30.8	11.20	43.9	15.96
19	Limestone[a]	500	3.5	ST	94.4	189	25.8	9.40	36.4	13.25
20	Limestone	1,000	0.7	ST	109.2	109	28.3	5.15	39.2	7.13
21	Limestone	1,000	2.0	ST	151.7	152	41.1	7.47	57.7	10.50
22	Limestone	1,000	3.5	ST	177.2	177	49.3	8.97	69.9	12.71
23	Limestone	1,000	5.0	ST	187.0	187	53.0	9.63	75.5	13.73

Note: No additives, landfill disposal in all cases.

[a] No reheat.

TABLE 27. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR LIMESTONE PROCESSES WITH FORCED OXIDATION AND ADDITIVES

Case	Absorbent	Plant size, MW	Coal sulfur, %	Absorber type	Capital investment, 1982$ M$	Capital investment, 1982$ $/kW	Annual revenue requirements, 1984$ First-year M$	Annual revenue requirements, 1984$ First-year Mills/kWh	Annual revenue requirements, 1984$ Levelized M$	Annual revenue requirements, 1984$ Levelized Mills/kWh
Adipic Acid										
24	Limestone	200	3.5	ST	54.4	272	15.3	13.95	21.9	19.86
25	Limestone	500	0.7	ST	61.4	123	16.4	5.97	23.0	8.36
26	Limestone	500	2.0	ST	79.1	158	21.6	7.86	30.4	11.07
27	Limestone	500	3.5	ST	92.8	186	25.8	9.38	36.6	13.31
28	Limestone	500	5.0	ST	97.9	196	27.7	10.07	39.5	14.36
29	Limestone	500	3.5	TCA	90.4	181	25.4	9.23	36.1	13.12
30	Limestone	500	3.5	VST	96.3	193	27.0	9.81	38.3	13.93
31	Limestone[a]	500	3.5	ST	100.7	201	26.8	9.73	37.3	13.58
32	Limestone[b]	500	3.5	ST	92.8	186	25.7	9.35	36.4	13.25
33	Limestone	1,000	3.5	ST	155.5	155	42.8	7.78	60.4	10.99
Magnesium Oxide										
34	Limestone	500	0.7	ST	69.1	138	18.4	6.70	25.8	9.37
35	Limestone	500	2.0	ST	85.9	172	24.0	8.72	34.1	12.39
36	Limestone	500	3.5	ST	102.5	205	28.9	10.49	41.1	14.93
37	Limestone	500	5.0	ST	107.9	216	30.8	11.18	43.9	15.98
38	Limestone	500	3.5	TCA	98.2	196	27.8	10.09	39.5	14.38
39	Limestone	500	3.5	VST	113.1	226	31.6	11.50	44.9	16.33
40	Limestone[a]	500	3.5	ST	111.2	222	28.5	10.36	39.2	14.27

Note: Landfill disposal except as noted.

a. Pond disposal.

b. Dibasic acid.

TABLE 28. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR LIME PROCESSES

Case	Absorbent	Plant size, MW	Coal sulfur, %	Absorber type	Capital investment, 1982$		Annual revenue requirements, 1984$			
							First-year		Levelized	
					M$	$/kW	M$	Mills/kWh	M$	Mills/kWh
Without Forced Oxidation (Pond Disposal)										
41	Lime	500	0.7	ST	60.7	121	15.3	5.56	20.9	7.62
42	Lime	500	2.0	ST	79.3	159	21.8	7.92	30.7	11.18
43	Lime	500	3.5	ST	96.2	192	28.0	10.19	40.3	14.66
44	Lime	500	5.0	ST	105.3	211	31.6	11.51	46.0	16.71
45	Lime	500	3.5	TCA	94.6	189	27.7	10.09	40.0	14.55
46	Lime	500	3.5	VST	103.2	206	29.9	10.88	43.0	15.63
With Forced Oxidation (Landfill Disposal)										
47	Lime	500	3.5	ST	88.9	178	28.5	10.35	42.1	15.31
48	Lime	500	3.5	TCA	86.2	172	27.9	10.16	41.5	15.07
49	Lime	500	3.5	VST	92.7	185	29.6	10.77	43.8	15.92
With Forced Oxidation and Magnesium Oxide (Landfill Disposal)										
50	Lime	500	3.5	ST	89.6	179	28.6	10.40	42.3	15.37
51	Lime	500	3.5	TCA	90.4	181	28.7	10.45	42.4	15.43
52	Lime	500	3.5	VST	96.8	194	30.3	11.01	44.5	16.18

TABLE 29. CAPITAL INVESTMENT COMPARISON OF PROCESSES WITH AND WITHOUT FORCED OXIDATION

		Capital investment, 1982 $/kW											
Case	Process[a], disposal	Materials handling	Feed preparation	Gas handling	SO_2 absorption	Oxidation	Reheat	Solids separation	Services, etc.	Disposal area	Total direct	Total indirect	Total
	0.7% Sulfur Coal												
2	No forced oxidation, pond	2.5	4.7	22.3	36.0	-	-	3.4	4.1	8.2	81.2	65.3	146.5
13	No forced oxidation, landfill	2.5	4.7	22.2	36.0	-	-	4.5	4.2	3.1	77.2	63.5	140.7
	2.0% Sulfur Coal												
3	No forced oxidation, pond	2.6	6.3	21.7	35.5	-	7.2	4.4	4.7	18.5	100.9	79.9	180.8
14	Forced oxidation, landfill	2.6	6.3	21.7	35.5	4.5	7.2	6.0	5.0	6.1	95.0	78.1	173.1
	3.5% Sulfur Coal												
4	No forced oxidation, pond	5.1	9.4	22.6	40.7	-	7.3	5.3	5.4	30.6	126.4	98.7	225.1
15	Forced oxidation, landfill	5.1	9.4	22.6	40.7	5.2	7.3	7.4	5.8	9.8	113.3	92.8	206.1
	5.0% Sulfur Coal												
5	No forced oxidation, pond	5.7	10.7	22.6	41.1	-	7.2	5.9	5.6	40.3	139.1	107.6	246.7
16	Forced oxidation, landfill	5.7	10.7	22.6	41.1	6.0	7.2	8.3	6.0	12.4	120.0	98.3	218.3

[a] 500-MW, limestone spray tower processes, in-loop forced oxidation for cases 14, 15, and 16, natural oxidation to 95% $CaSO_4 \cdot 2H_2O$ for cases 2 and 13.

TABLE 30. ANNUAL REVENUE REQUIREMENTS COMPARISON OF PROCESSES WITH AND WITHOUT FORCED OXIDATION

Case	Process[a], disposal	First-year annual revenue requirements, 1984 mills/kWh: Absorbent	Labor	Steam	Water	Electricity	Diesel fuel	Maintenance[b]	Overheads	Capital charges	Total	Levelized annual revenue requirements, 1984 mills/kWh
	0.7% Sulfur Coal											
2	No forced oxidation, pond	0.07	0.14	-	0.01	0.53	-	1.13	0.77	3.91	6.56	8.91
13	No forced oxidation, landfill	0.07	0.41	-	0.01	0.53	0.02	1.12	0.92	3.76	6.84	9.56
	2.0% Sulfur Coal											
3	No forced oxidation, pond	0.23	0.17	0.50	0.01	0.48	-	1.34	0.90	4.83	8.46	11.68
14	Forced oxidation, landfill	0.23	0.43	0.50	0.01	0.53	0.03	1.37	1.08	4.63	8.80	12.49
	3.5% Sulfur Coal											
4	No forced oxidation, pond	0.44	0.17	0.49	0.01	0.68	-	1.60	1.06	6.02	10.47	14.42
15	Forced oxidation, landfill	0.44	0.46	0.49	0.01	0.78	0.06	1.60	1.23	5.51	10.58	15.08
	5.0% Sulfur Coal											
5	No forced oxidation, pond	0.63	0.17	0.49	0.02	0.73	-	1.69	1.12	6.59	11.44	15.73
16	Forced oxidation, landfill	0.63	0.49	0.49	0.01	0.86	0.07	1.67	1.30	5.84	11.36	16.25

a. 500-MW, limestone spray tower processes, in-loop forced oxidation for cases 14, 15, and 16, natural oxidation to 95% $CaSO_4 \cdot 2H_2O$ for cases 2 and 13.

b. Includes analysis.

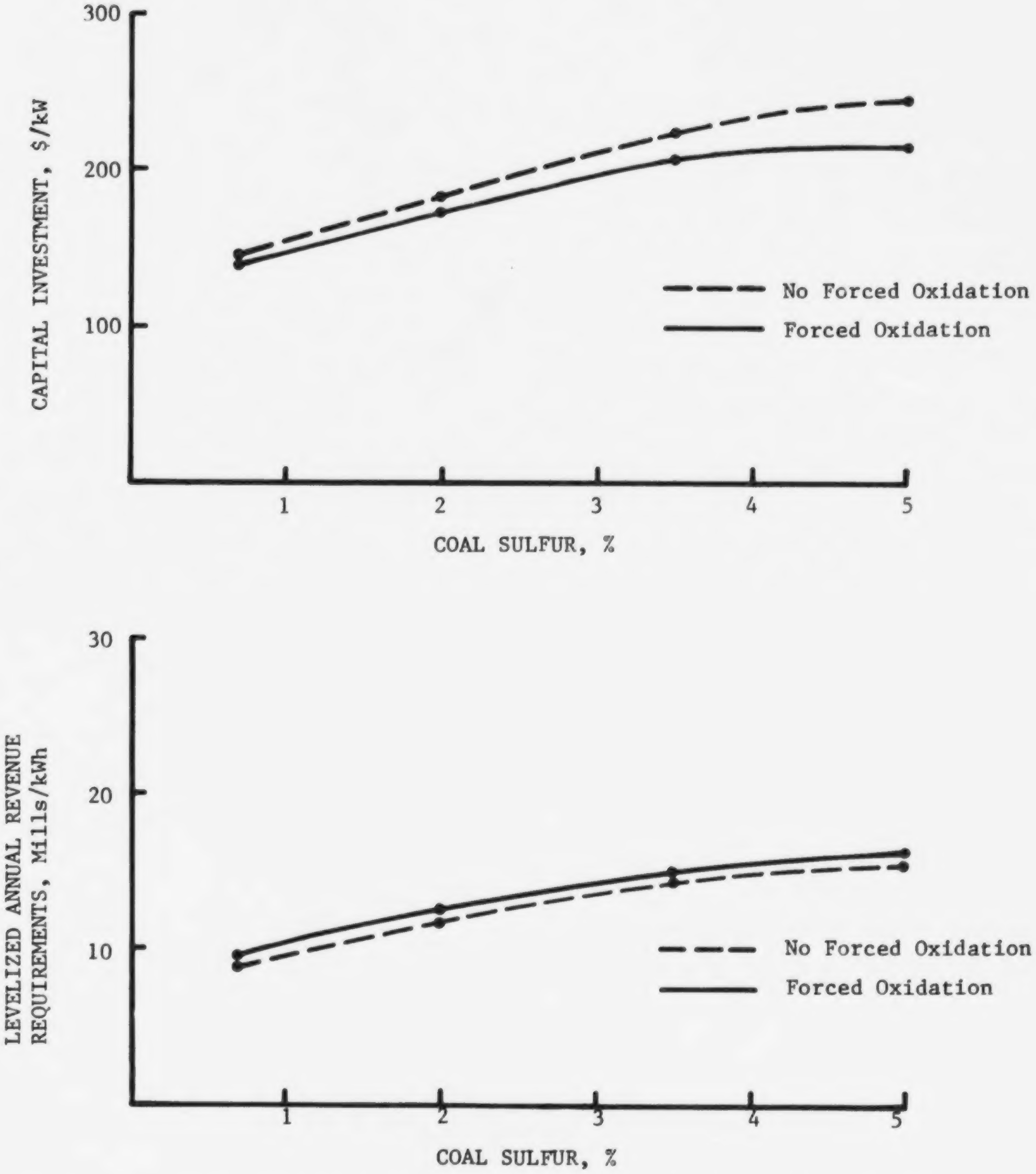


Figure 14. Cost comparison of 500-MW limestone processes with and without forced oxidation at different coal sulfur contents.

Except for the 5.0% sulfur coal cases, first-year annual revenue requirements are higher for the forced-oxidation cases. Even though the forced-oxidation processes have lower capital charges in all cases, this cost advantage is offset at the lower coal sulfur levels by the higher operating labor and overhead costs associated with the waste dewatering and trucking operations. In the 5.0% sulfur coal case the lower capital charges for the forced-oxidation process (because of the lower disposal area investment) is sufficient to compensate for the higher labor and overhead costs.

In contrast to the first-year annual revenue requirements, however, levelized annual revenue requirements are lower for the processes without forced oxidation in all cases. The lower levelized annual revenue requirements for the processes without forced oxidation are a result of the smaller proportion of operating and maintenance costs in the total annual revenue requirements of these processes. To calculate levelized annual revenue requirements, the operating and maintenance costs of the first-year annual revenue requirements are multiplied by the levelizing factor of 1.886 and added to the other component of annual revenue requirements, the capital charges. (The capital charges of both first-year and levelized annual revenue requirements are levelized and do not differ.) Therefore, the levelized annual revenue requirements of processes with a low proportion of operating and maintenance costs and a high proportion of capital charges differ less from first-year annual revenue requirements than do those for which the reverse is true. Processes without forced oxidation have low operating costs and high capital investments (due to pond construction costs) compared with the more complex processes with landfill disposal. The conclusions reached when comparing such processes may differ, depending on which type of annual revenue requirements is used. Assuming the yearly 6% inflation and 10% cost of money used in the levelizing factor to be valid over the life of the installation, the processes without forced oxidation would have lower lifetime annual revenue requirements than the forced-oxidation processes. For the 3.5% sulfur coal, 500-MW case, the lifetime difference is about $54,000,000 (0.66 mills/kWh).

Comparison of Limestone Processes With and Without Additives

A comparison of costs by process area for limestone processes without additives and with magnesium oxide or adipic acid additive is shown in Tables 31 and 32. For the forced-oxidation cases the results are shown for various coal sulfur levels while for the cases without forced oxidation the results are shown for only the 3.5% sulfur coal. The capital investments and levelized annual revenue requirements for the forced-oxidation processes with and without additives as a function of the coal sulfur level are shown in Figure 15.

Magnesium Oxide--

The use of magnesium oxide has no significant effect (less than a 2.5% reduction) on capital investment. The equipment needed to handle the relatively small volumes of magnesium oxide does not materially affect costs. By improving SO_2 removal efficiency and absorbent utilization the use of this additive slightly decreases gas handling, SO_2 absorption, and disposal

TABLE 31. CAPITAL INVESTMENT COMPARISON FOR PROCESSES WITH AND WITHOUT ADDITIVES

		Capital investment, 1982 \$/kW											
Case		Materials handling	Feed preparation	Gas handling	SO_2 absorption	Oxidation	Reheat	Solids separation	Services, etc.	Disposal area	Total direct	Total indirect	Total
	Landfill Disposal[a]												
	0.7% sulfur coal												
13	No additive	2.5	4.7	22.2	36.0	-	-	4.5	4.2	3.1	77.2	63.5	140.7
25	Adipic acid	2.1	3.9	20.7	30.3	-	-	3.9	3.8	2.8	67.5	55.3	122.8
34	MgO	2.6	4.8	22.0	34.8	-	-	4.6	4.1	3.0	75.9	62.3	138.2
	2.0% sulfur coal												
14	No additive	2.6	6.3	21.7	35.5	4.5	7.2	6.0	5.0	6.1	95.0	78.1	173.1
26	Adipic acid	2.7	5.7	19.6	33.2	4.7	5.2	5.7	4.6	5.5	86.9	71.3	158.2
35	MgO	2.7	6.2	21.5	34.7	5.0	7.2	6.0	5.0	6.0	94.3	77.5	171.8
	3.5% sulfur coal												
15	No additive	5.1	9.4	22.6	40.7	5.2	7.3	7.4	5.8	9.8	113.3	92.8	206.1
27	Adipic acid	5.1	8.6	21.0	35.4	4.7	6.4	6.9	5.3	8.5	101.9	83.6	185.5
36	MgO	5.1	9.4	22.4	39.8	6.2	7.3	7.4	5.8	9.2	112.6	92.4	205.0
	5.0% sulfur coal												
16	No additive	5.7	10.7	22.6	41.1	6.0	7.2	8.3	6.0	12.4	120.0	98.3	218.3
28	Adipic acid	5.3	9.9	21.0	35.6	5.2	6.5	7.8	5.5	11.1	107.7	88.2	195.9
37	MgO	5.4	10.3	22.3	39.1	7.8	7.2	8.5	5.9	12.1	118.6	97.2	215.8
	Pond Disposal[b]												
	3.5% sulfur coal												
4	No additive	5.1	9.4	22.6	40.7	-	7.3	5.3	5.4	30.6	126.4	98.7	225.1
31	Adipic acid	5.2	8.6	21.0	35.5	-	6.5	4.9	5.0	26.2	112.9	88.5	201.4
40	MgO	5.1	9.4	22.6	40.5	-	7.3	5.2	5.4	29.3	124.8	97.6	222.4

a. 500-MW, limestone spray tower processes, in-loop forced oxidation for no additives and adipic acid cases except as noted, bleedstream forced oxidation for all MgO processes, natural oxidation for all 0.7% sulfur coal cases.

b. 500-MW, limestone spray tower processes without forced oxidation.

TABLE 32. ANNUAL REVENUE REQUIREMENTS COMPARISON

FOR PROCESSES WITH AND WITHOUT ADDITIVES

		First-year annual revenue requirements, 1984 mills/kWh											Levelized annual revenue requirements, 1984 mills/kWh
Case		Absorbent	Additive	Labor	Steam	Water	Electricity	Diesel fuel	Maintenance[a]	Overheads	Capital charges	Total	
	Landfill Disposal[b]												
	0.7% sulfur coal												
13	No additive	0.07	-	0.41	-	0.01	0.53	0.02	1.12	0.92	3.76	6.84	9.56
25	Adipic acid	0.05	0.01	0.40	-	0.01	0.39	0.02	0.98	0.83	3.28	5.97	8.36
34	MgO	0.06	0.00	0.41	-	0.01	0.50	0.02	1.10	0.91	3.69	6.70	9.37
	2.0% sulfur coal												
14	No additive	0.23	-	0.43	0.50	0.01	0.53	0.03	1.37	1.08	4.63	8.80	12.49
26	Adipic acid	0.18	0.04	0.42	0.24	0.01	0.48	0.03	1.24	0.99	4.23	7.86	11.07
35	MgO	0.21	0.01	0.43	0.50	0.01	0.51	0.03	1.36	1.07	4.59	8.72	12.39
	3.5% sulfur coal												
15	No additive	0.44	-	0.46	0.49	0.01	0.78	0.06	1.60	1.23	5.51	10.58	15.08
27	Adipic acid	0.34	0.08	0.42	0.39	0.01	0.61	0.04	1.43	1.11	4.95	9.38	13.31
36	MgO	0.41	0.02	0.46	0.49	0.01	0.75	0.04	1.59	1.24	5.48	10.49	14.93
	5.0% sulfur coal												
16	No additive	0.63	-	0.49	0.49	0.01	0.86	0.07	1.67	1.30	5.84	11.36	16.25
28	Adipic acid	0.48	0.11	0.46	0.39	0.01	0.67	0.06	1.49	1.16	5.24	10.07	14.36
37	MgO	0.59	0.03	0.49	0.49	0.01	0.80	0.07	1.65	1.29	5.76	11.18	15.98
	Pond Disposal[c]												
	3.5% sulfur coal												
4	No additive	0.44	-	0.17	0.49	0.01	0.68	-	1.60	1.06	6.02	10.47	14.42
31	Adipic acid	0.34	0.54	0.17	0.39	0.01	0.52	-	1.43	0.95	5.38	9.73	13.58
40	MgO	0.41	0.02	0.17	0.49	0.01	0.67	-	1.59	1.05	5.95	10.36	14.27

a. Includes analysis.

b. 500-MW, limestone spray tower processes, in-loop forced oxidation for no additives and adipic acid cases except as noted, bleedstream forced oxidation for all MgO processes, natural oxidation for all 0.7% sulfur coal cases.

c. 500-MW, limestone spray tower processes without forced oxidation.

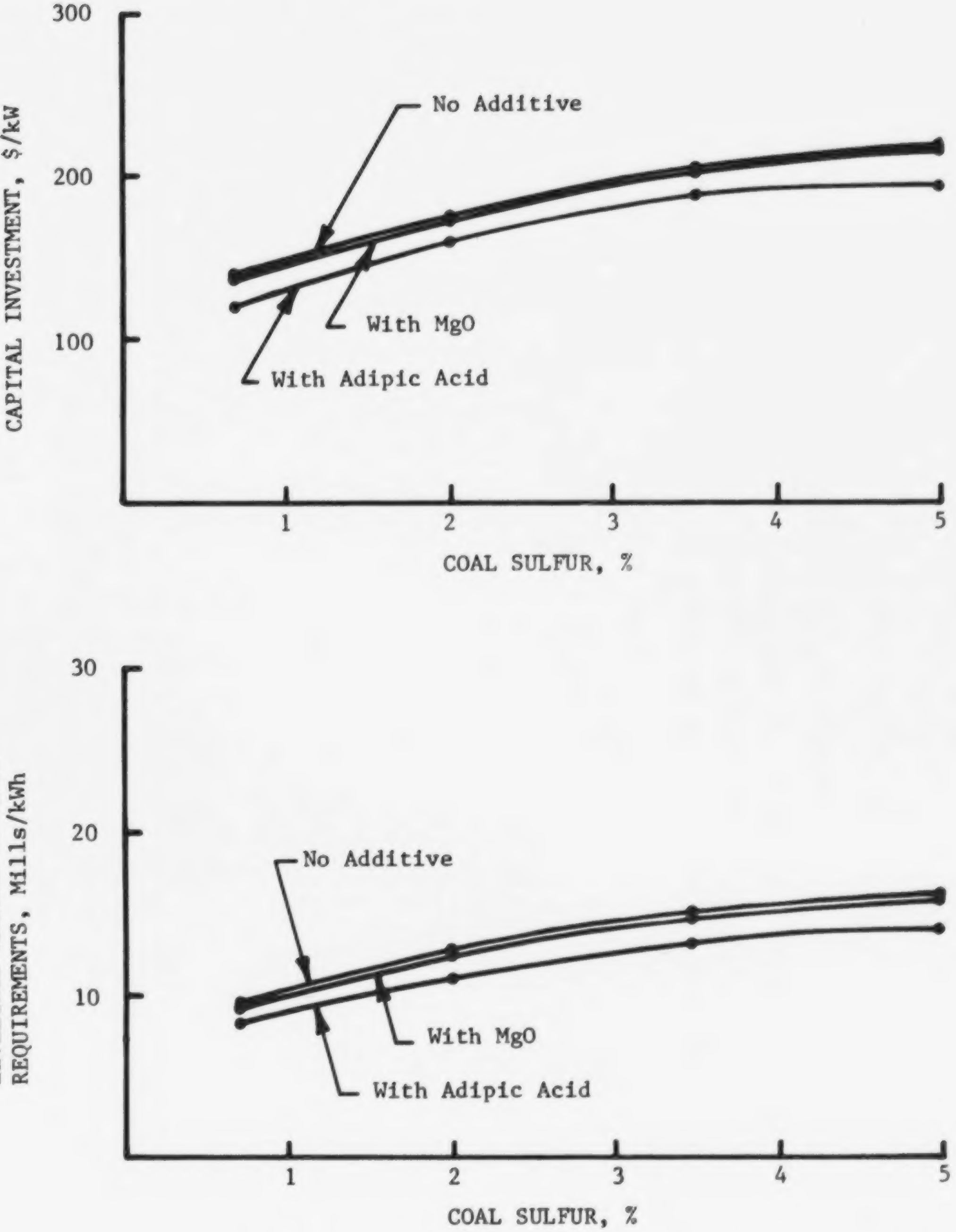


Figure 15. Cost comparison of limestone forced-oxidation processes with and without additives at different coal sulfur contents.

costs. This effect is thus more pronounced at high coal sulfur contents. The increased equipment required for the magnesium oxide addition, although somewhat offset by the reduced absorbent consumption, still results in a marginal increase in the material handling and feed preparation areas for the low-sulfur coal cases. For higher sulfur coal cases the increased absorbent utilization results in lower costs in these areas. The necessity of using bleedstream oxidation results in a higher oxidation area cost for the higher sulfur coal cases.

The effect of magnesium oxide addition on first-year annual revenue requirements of forced-oxidation processes is similar to the effect on capital investment. The cost of the additive is minimal (less than 1% of total annual revenue requirements). The improved SO_2 removal efficiency and absorbent utilization, expressed primarily as reduced absorbent and electricity costs, more than offset the cost for magnesium oxide in all cases. (In fact, the saving in absorbent alone more than compensates for the cost of magnesium oxide.) For the 3.5% sulfur coal case, the cost reduction associated with using magnesium oxide, however, is less than 1%. For the case without forced oxidation where more of the dissolved portion of the additive is lost in the low-solids waste to the pond, the effect on first-year annual revenue requirements is relatively minor, decreasing costs only about 1%. In terms of levelized annual revenue requirements the results are essentially the same; i.e., the use of magnesium oxide as an additive decreases the annual revenue requirements but the decrease is relatively insignificant.

Adipic Acid--

In contrast to the use of magnesium oxide, the use of adipic acid appears to be an economically attractive additive for all of the applications considered in this study. Capital investment reductions range from nearly 13% for the 0.7% sulfur western subbituminous coal case to about 10% for the high-sulfur coal cases. As in the magnesium oxide addition cases, the additional equipment needed to handle the small quantities of adipic acid does not have an appreciable effect on equipment costs in the material handling and feed preparation areas. However, by improving SO_2 removal efficiency and absorbent utilization, the use of this additive decreases the capital investments for the feed preparation, gas handling, SO_2 absorption, solids separation, and disposal area costs. The reason for the more dramatic reductions in costs when using adipic acid is the assumption that with this additive the limestone stoichiometry can be reduced by one-fourth (1.07 versus 1.40).

The use of adipic acid reduces both first-year and levelized annual revenue requirements for forced-oxidation processes by 11% to 13% depending on the coal sulfur level. In all forced-oxidation cases the cost of adipic acid is minimal (less than 1% of annual revenue requirements for all except the 5.0% sulfur coal case). The improved SO_2 removal efficiency and absorbent utilization, expressed primarily as reduced capital charges, overheads and maintenance, absorbent, and electricity costs, more than offset the cost for adipic acid in all cases. (As with magnesium oxide, the saving in absorbent more than compensates for the adipic acid cost.) For the cases without forced oxidation where more of the additive is lost in the low-solids waste, the resulting higher adipic acid costs decrease the savings in annual revenue requirements. For the 3.5% sulfur coal the reduction is 11% for the forced-oxidation case but only about 7% for the case without forced oxidation and the adipic acid cost accounts for 5.5% of the annual revenue requirements.

It is apparent that adipic acid is a more cost-effective additive than magnesium oxide. It is also apparent that adipic acid addition seems to be an economically attractive alternative to conventional, unenhanced limestone scrubbing. Cost differences of these magnitudes would not normally be regarded as highly significant in cost studies. The nature of these comparisons, however, allows more decisive conclusions. The processes are very similar except for the use of additives. Excluding extreme variations in adipic acid cost (relative to other cost variations) the fairly well-documented improvements in absorbent utilization and SO_2 removal efficiency indicate that adipic acid addition provides a real cost advantage for limestone scrubbing processes, particularly those incorporating high levels of waste dewatering and water reuse. In the case of direct pond disposal the cost advantages of processes using additives are reduced because of the additive lost in the low-solids wastes. The cost of the additive in these cases is approximately the same as the limestone cost.

Power Unit Size and Sulfur Contents

Cost comparisons by process area for different power unit sizes are shown in Tables 33 and 34. The effect of power unit size on capital investment and levelized annual revenue requirements at different coal sulfur contents is shown in Figure 16 for the forced-oxidation cases. The usual effect of economies of scale in terms of unit costs ($/kW, mills/kWh) is evident for all conditions. For a given coal sulfur level the capital investment of 500-MW cases is about 33% less expensive, in terms of $/kW, than the 200-MW cases and the 1,000-MW cases are about 13% less expensive than the 500-MW cases. The same effects are seen in the annual revenue requirements. The effects are also essentially the same for the processes without forced oxidation.

Comparisons by Type of Absorber

Cost comparisons by process area for limestone processes with different types of absorbers are shown in Tables 35 and 36. The comparisons are for limestone processes at the 500-MW, 3.5% sulfur coal conditions. The capital investment relationships of the processes with TCA's, spray towers, and venturi - spray towers remain the same whether forced oxidation or additives are used. In all cases the TCA process has the lowest capital investment and the venturi - spray tower has the highest.

The major capital investment advantage of the TCA process lies in lower costs for the SO_2 absorption area because of the smaller absorber and lower L/G. Capital investment for the reheat area is also slightly lower because of the higher pressure drop in the TCA, and subsequent higher heat compression through the ID fan. Processes using the larger, lower pressure drop spray tower have higher capital investments for the SO_2 absorption and reheat

TABLE 33. CAPITAL INVESTMENT COMPARISON AT VARIOUS POWER UNIT SIZES

		Capital investment, 1982 $/kW											
Case	Process[a]	Materials handling	Feed preparation	Gas handling	SO_2 absorption	Oxidation	Reheat	Solids separation	Services, etc.	Disposal area	Total direct	Total indirect	Total
200 MW													
1	No forced oxidation	6.4	13.7	30.0	54.0	-	9.7	10.3	7.5	38.6	170.2	143.4	313.6
11	Forced oxidation	6.4	13.7	29.9	53.9	7.6	9.7	13.9	8.2	13.3	156.6	138.7	295.3
24	Forced oxidation, adipic acid	6.9	13.9	27.8	47.3	6.8	8.6	13.4	7.5	11.9	144.1	127.8	271.9
500 MW													
4	No forced oxidation	5.1	9.4	22.6	40.7	-	7.3	5.3	5.4	30.6	126.4	98.7	225.1
15	Forced oxidation	5.1	9.4	22.6	40.7	5.2	7.3	7.4	5.8	9.8	113.3	92.8	206.1
27	Forced oxidation, adipic acid	5.1	8.6	21.0	35.4	4.7	6.4	6.9	5.3	8.5	101.9	83.6	185.5
1,000 MW													
8	No forced oxidation	2.9	6.0	21.8	39.9	-	7.2	3.3	4.9	25.4	111.4	80.7	192.1
22	Forced oxidation	2.9	6.0	21.7	39.9	5.5	7.1	4.7	5.4	7.9	101.1	76.1	177.2
33	Forced oxidation, adipic acid	2.9	5.6	19.9	34.0	4.4	6.2	4.4	4.5	6.7	88.6	66.9	155.5

a. 3.5% sulfur coal, limestone spray tower processes.

TABLE 34. ANNUAL REVENUE REQUIREMENTS COMPARISON AT VARIOUS POWER UNIT SIZES

		First-year annual revenue requirements, 1984 mills/kWh											Levelized annual revenue requirements, 1984 mills/kWh
Case	Process[a]	Absorbent	Additive	Labor	Steam	Water	Electricity	Diesel fuel	Maintenance[b]	Overheads	Capital charges	Total	
200 MW													
1	No forced oxidation	0.45	-	0.28	0.50	0.01	0.72	-	2.43	1.62	8.38	14.39	19.72
11	Forced oxidation	0.45	-	0.93	0.50	0.01	0.82	0.08	2.48	2.05	7.89	15.21	21.69
24	Forced oxidation, adipic acid	0.34	0.08	0.92	0.39	0.01	0.65	0.06	2.29	1.94	7.27	13.95	19.86
500 MW													
4	No forced oxidation	0.44	-	0.17	0.49	0.01	0.68	-	1.60	1.06	6.02	10.47	14.42
15	Forced oxidation	0.44	-	0.46	0.49	0.01	0.78	0.06	1.60	1.23	5.51	10.58	15.08
27	Forced oxidation, adipic acid	0.34	0.08	0.42	0.39	0.01	0.61	0.04	1.43	1.11	4.95	9.38	13.31
1,000 MW													
8	No forced oxidation	0.43	-	0.11	0.48	0.01	0.65	-	1.26	0.82	5.13	8.89	12.23
22	Forced oxidation	0.43	-	0.33	0.48	0.01	0.73	0.05	1.25	0.95	4.74	8.97	12.71
33	Forced oxidation, adipic acid	0.33	0.07	0.29	0.37	0.01	0.57	0.04	1.11	0.84	4.15	7.78	10.99

a. 3.5% sulfur coal, limestone spray tower processes.
b. Includes analysis.

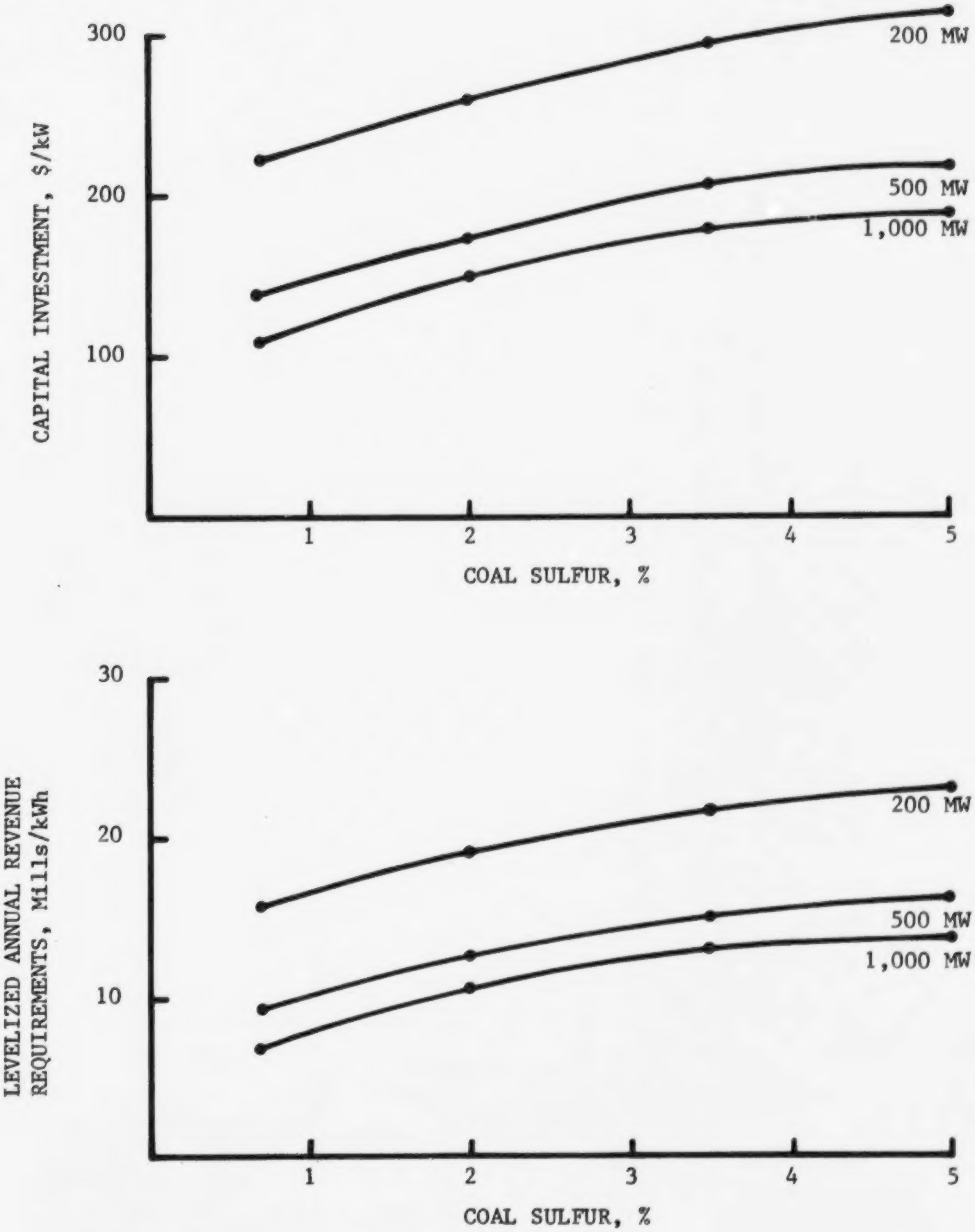


Figure 16. Cost comparison of power unit size and coal sulfur content for limestone forced-oxidation processes.

TABLE 35. CAPITAL INVESTMENT COMPARISON BY ABSORBER AND PROCESS TYPE

Case	Process[a]	Capital investment, 1982 $/kW											
		Materials handling	Feed preparation	Gas handling	SO_2 absorption	Oxidation	Reheat	Solids separation	Services, etc.	Disposal area	Total direct	Total indirect	Total
	Without Forced Oxidation (Pond)												
4	Spray tower	5.1	9.4	22.6	40.7	-	7.3	5.3	5.4	30.6	126.4	98.7	225.1
6	TCA	5.1	9.5	24.2	34.0	-	7.0	5.3	5.1	31.0	121.2	94.4	215.6
7	Venturi - spray tower	5.0	9.2	26.8	47.6	-	7.1	5.2	6.1	29.3	136.3	107.1	243.4
	With Forced Oxidation (Landfill)												
15	Spray tower	5.1	9.4	22.6	40.7	5.2	7.3	7.4	5.8	9.8	113.3	92.8	206.1
17	TCA	5.1	9.5	23.0	36.5	4.3	7.1	7.3	5.6	9.9	108.3	88.7	197.0
18	Venturi - spray tower	5.0	9.3	26.7	46.5	2.7	7.1	7.2	6.2	9.2	119.9	98.5	218.4
	With Forced Oxidation and Adipic Acid (Landfill)												
27	Spray tower	5.1	8.6	21.0	35.5	4.7	6.4	6.9	5.2	8.5	101.9	83.6	185.5
29	TCA	5.1	9.2	20.9	32.8	4.1	6.3	7.0	5.1	8.9	99.4	81.5	180.9
30	Venturi - spray tower	5.1	9.2	24.0	37.2	2.8	6.2	7.0	5.4	8.9	105.8	86.9	192.7

a. 500-MW, 3.5% sulfur coal, limestone process.

TABLE 36. ANNUAL REVENUE REQUIREMENTS COMPARISON BY ABSORBER AND PROCESS TYPE

		First-year annual revenue requirements, 1984 mills/kWh											Levelized annual revenue requirements, 1984 mills/kWh
		Raw materials											
Case	Process[a]	Absorbent	Additive	Labor	Steam	Water	Electricity	Diesel fuel	Maintenance[b]	Overheads	Capital charges	Total	
	Without Forced Oxidation (Pond)												
4	Spray tower	0.44	-	0.17	0.49	0.01	0.68	-	1.60	1.06	6.02	10.47	14.42
6	TCA	0.45	-	0.17	0.46	0.01	0.65	-	1.52	1.01	5.76	10.03	13.81
7	Venturi - spray tower	0.41	-	0.18	0.45	0.01	0.91	-	1.75	1.16	6.50	11.37	15.68
	With Forced Oxidation (Landfill)												
15	Spray tower	0.44	-	0.46	0.49	0.01	0.78	0.06	1.60	1.23	5.51	10.58	15.08
17	TCA	0.45	-	0.46	0.48	0.01	0.65	0.06	1.53	1.19	5.26	10.09	14.37
18	Venturi - spray tower	0.41	-	0.47	0.45	0.01	0.98	0.04	1.70	1.30	5.84	11.20	15.96
	With Forced Oxidation and Adipic Acid (Landfill)												
27	Spray tower	0.34	0.08	0.42	0.39	0.01	0.61	0.04	1.43	1.10	4.96	9.38	13.31
29	TCA	0.38	0.10	0.46	0.37	0.01	0.53	0.04	1.39	1.12	4.83	9.23	13.12
30	Venturi - spray tower	0.38	0.05	0.47	0.35	0.01	0.71	0.04	1.48	1.17	5.15	9.81	13.93

a. 500-MW, 3.5% sulfur coal, limestone process.
b. Includes analysis.

areas than the TCA processes. This is only partially offset by the lower capital investment in the gas handling area due to the lower ID fan requirements. The processes using the venturi - spray tower absorbers have a higher limestone utilization rate, and thus slightly lower capital investments for the feed preparation, solids disposal (dewatering), and disposal areas. This is more than offset by the higher capital investments for the gas handling and SO_2 absorption areas, however, making processes using this absorber combination about one-tenth higher in capital investment than processes using a TCA or spray tower.

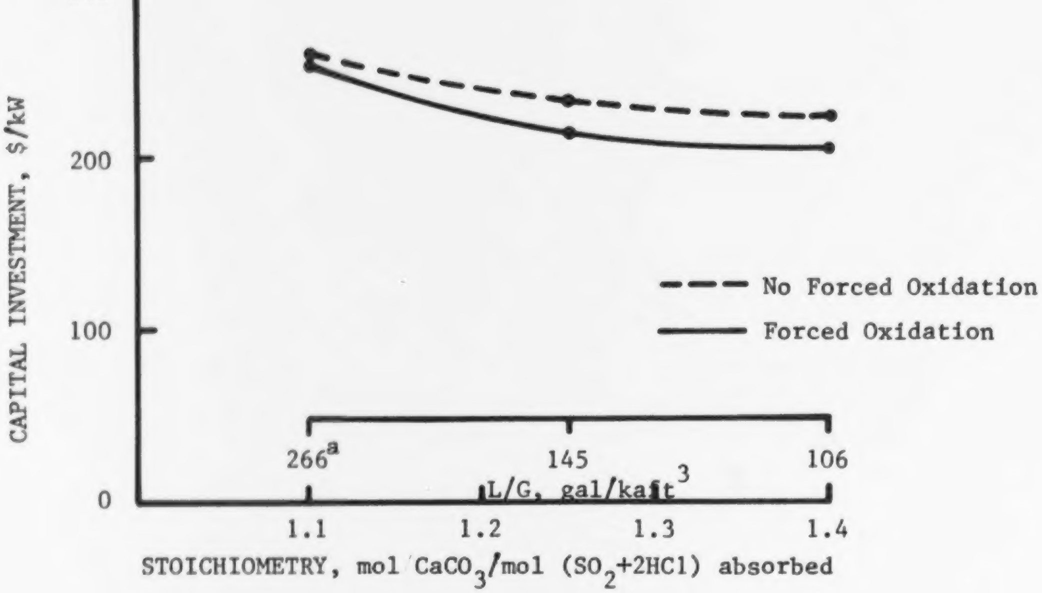
The same cost relationships also prevail in first-year and levelized annual revenue requirements. Processes using the TCA have the lowest annual revenue requirements and processes using the venturi - spray tower have the highest. In this case, the differences are largely the result of costs based on capital investment such as maintenance and capital charges. In conversion costs other than maintenance there is little difference in annual revenue requirements between the processes using TCA's, spray towers, or venturi - spray towers. (The role of maintenance costs in the economic relationship of processes with TCA's and processes with spray towers is discussed in the section on cost sensitivities.) Processes with venturi - spray towers have slightly lower absorbent costs but about one-fifth higher electricity costs. Processes with spray towers have higher reheat steam costs. Other conversion costs do not differ greatly.

Spray Tower Absorber Design Conditions

Since the superficial gas velocity through the absorber and the SO_2 removal efficiency are normally determined by the technology or the performance requirements, the absorber L/G and stoichiometry are the only remaining design conditions that can be optimized. Thus it becomes an economic tradeoff between the increased costs associated with larger slurry recirculation pumps (L/G) and the costs for the limestone preparation equipment and the waste disposal area (stoichiometry). As can be seen from Figure 17, increasing the limestone stoichiometry and decreasing the absorber L/G is the more economically attractive option. Increasing the stoichiometry from 1.1 to 1.4 results in a decrease in capital investment of 18% for the forced-oxidation process and nearly 14% for the process without forced oxidation. Levelized annual revenue requirements decrease 18% and more than 16% respectively.

Partial Scrubbing Versus Full Scrubbing

For lower sulfur coal applications where relatively modest SO_2 removal efficiencies are required, two alternative scrubbing designs are possible: full scrubbing at the required removal efficiency or scrubbing part of the flue gas at a high removal efficiency (i.e., 90%) and bypassing the rest of the flue gas around the scrubber to reduce or eliminate the need for stack gas reheat. Both designs are technically feasible; the only questions are which is more economically attractive at low-sulfur applications and at what coal sulfur levels does the economic advantage of partial scrubbing disappear. Figure 18 compares the costs of full scrubbing and partial scrubbing for a



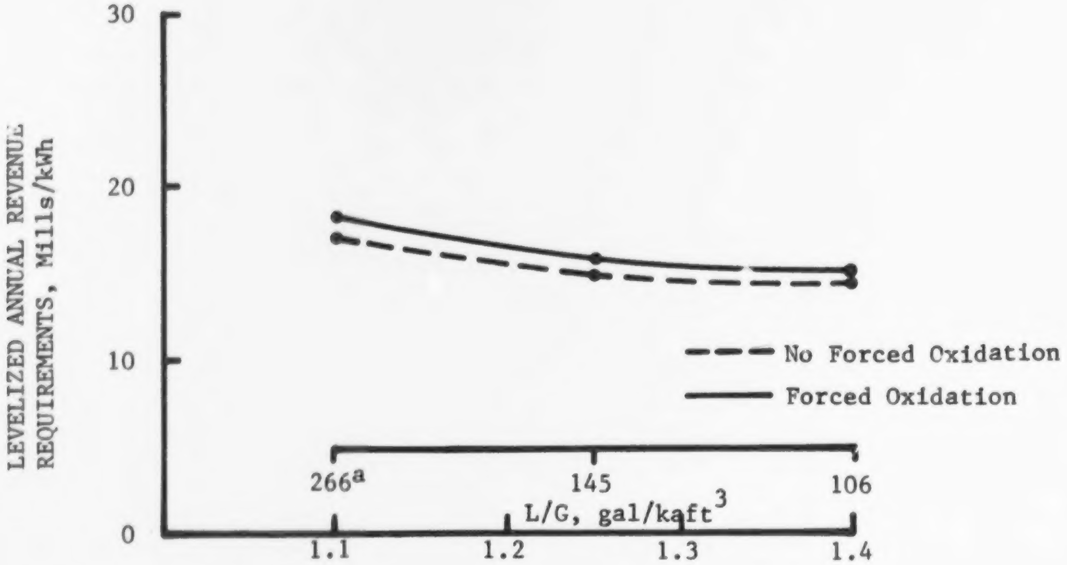


Figure 17. Cost comparison of limestone processes with and without forced oxidation at different absorber stoichiometries.

a. Extrapolation

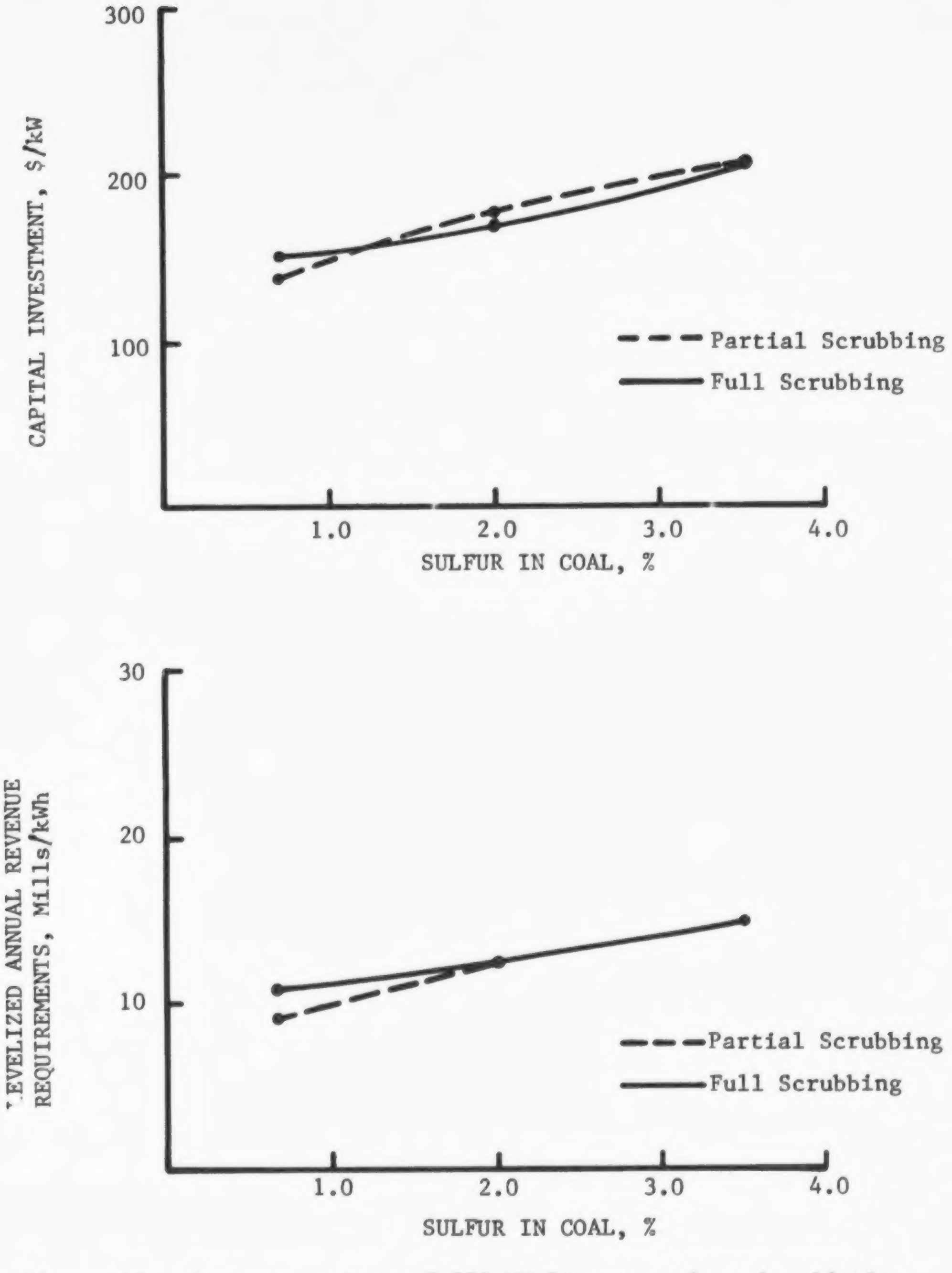


Figure 18. Cost comparison of 500-MW limestone forced-oxidation processes with partial or full scrubbing at different coal sulfur contents.

range of coal sulfur levels. These costs are based on a 500-MW limestone process with a spray tower absorber, forced oxidation, and landfill disposal. The partial scrubbing cases are based on 90% SO_2 removal in the absorber. As can be seen for the coal sulfur levels below 2%, partial scrubbing is more economical (in terms of levelized annual revenue requirements). In terms of the capital investment, partial scrubbing is less expensive for coals with sulfur contents below about 1.5%. For the 0.7% sulfur western subbituminous coal, partial scrubbing results in a saving of about $10/kW in capital investment (a 6.5% reduction) and about 1.2 mills/kWh in levelized annual revenue requirements (an 11.2% reduction). For the 2.0% sulfur coal, full scrubbing results in a $4.0/kW saving in capital investment over the partial scrubbing case but the levelized annual revenue requirements are essentially identical.

Full-Scrubbing Processes at 85%, 90%, and 95% SO_2 Removal

Another advantage of using adipic acid in a limestone scrubbing process is shown in Figure 19. The economics of a spray tower, forced-oxidation, and landfill disposal process with and without adipic acid are shown as a function of SO_2 removal efficiency. As the SO_2 removal efficiency increases from 85% to 95%, the economic advantage of the adipic-acid-enhanced process increases. At 85% removal the adipic-acid-enhanced process is about 9% lower in capital investment and nearly 12% lower in levelized annual revenue requirements. At 95% removal the corresponding reductions are more than 15% for both capital investment and levelized annual revenue requirements.

No Reheat

Since inline steam reheating involves significant expenditures in terms of both capital investment and annual revenue requirements and the benefits of reheating cannot be easily identified, some utilities have recently begun specifying FGD systems without reheat. Table 37 compares the economics of two limestone spray tower, forced-oxidation, and landfill disposal processes. The only difference between the two is that one has reheat and one does not. It should be noted that this comparison represents the most optimistic assumptions for the no-reheat case. In actual practice one might expect maintenance costs for the no-reheat case to be at least equal to, if not higher than, those for the reheat case.

The capital investment for the no-reheat case is more than 8% lower ($18/kW) than the reheat case, mostly due to the elimination of the reheater area. In first-year and levelized annual revenue requirements, the no-reheat case is about 11% and 12% lower respectively. These lower costs are due partly to the elimination of steam consumption and partly to the reduction in maintenance, overheads, and capital charges (these last reductions are due to the lower capital investment for the no-reheat case).

Comparisons of Limestone and Lime Processes

Limestone and lime process costs by process area are compared for different coal sulfur contents and by process type in Tables 38 and 39. The

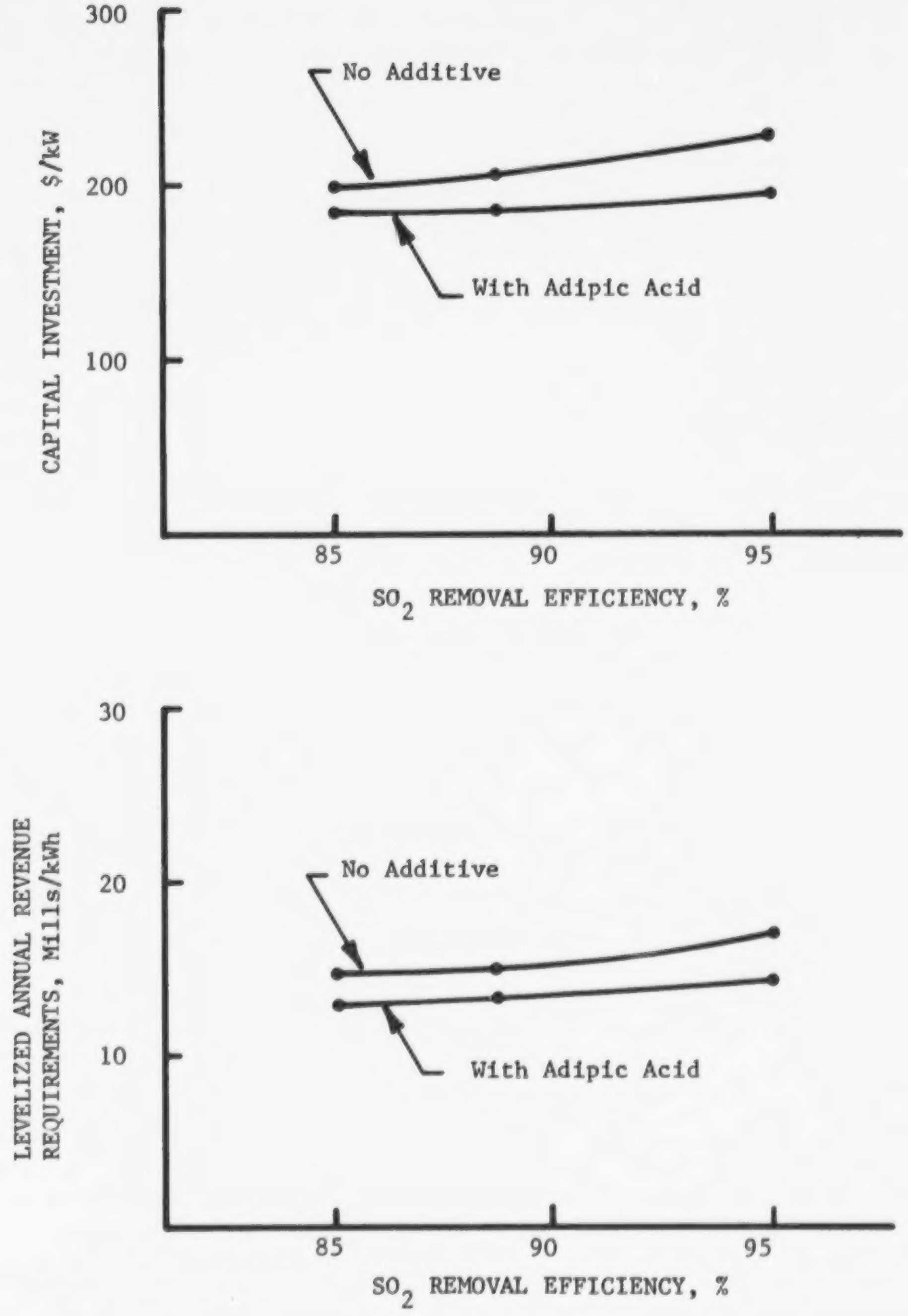


Figure 19. Cost comparison of 500-MW limestone forced-oxidation processes with and without adipic acid at different SO_2 removal efficiencies.

TABLE 37. CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS

FOR LIMESTONE PROCESSES WITH AND WITHOUT REHEAT

		Capital investment, 1982 $/kW											
Case	Process[a]	Materials handling	Feed preparation	Gas handling	SO_2 absorption	Oxidation	Reheat	Solids separation	Services, etc.	Disposal area	Total direct	Total indirect	Total
15	With reheat	5.1	9.4	22.6	40.7	5.2	7.3	7.4	5.8	9.8	113.3	92.8	206.1
19	Without reheat	5.1	9.4	22.2	39.4	5.3	-	7.4	5.3	9.8	103.9	85.0	188.9

		First-year annual revenue requirements, 1984 mills/kWh										Levelized annual revenue requirements, 1984 mills/kWh
Case	Process[a]	Absorbent	Labor	Steam	Water	Electricity	Diesel fuel	Maintenance[b]	Overheads	Capital charges	Total	
15	With reheat	0.44	0.46	0.49	0.01	0.78	0.06	1.60	1.23	5.51	10.58	15.08
19	Without reheat	0.44	0.46	-	0.01	0.76	0.06	1.46	1.16	5.05	9.40	13.25

a. 500-MW, 3.5% sulfur, limestone spray tower process with forced oxidation and landfill.

b. Includes analysis.

TABLE 38. CAPITAL INVESTMENT COMPARISON FOR LIMESTONE AND LIME PROCESSES

		Capital investment, 1982 $/kW											
Case	Process	Materials handling	Feed preparation	Gas handling	SO_2 absorption	Oxidation	Reheat	Solids separation	Services, etc.	Disposal area	Total direct	Total indirect	Total
	Spray Tower[a]												
	0.7% sulfur coal												
2	Limestone	2.5	4.7	22.3	36.0	-	-	3.4	4.1	8.2	81.2	65.3	146.5
41	Lime	2.3	1.9	20.7	28.8	-	-	3.2	3.4	7.0	67.3	54.2	121.5
	2.0% sulfur coal												
3	Limestone	2.6	6.3	21.7	35.5	-	7.2	4.4	4.7	18.5	100.9	79.9	180.8
42	Lime	3.8	2.8	19.6	32.5	-	5.9	4.1	4.1	15.5	88.3	70.3	158.6
	3.5% sulfur coal												
4	Limestone	5.1	9.4	22.6	40.7	-	7.3	5.3	5.4	30.6	126.4	98.7	225.1
43	Lime	4.8	3.4	21.4	35.8	-	7.1	4.9	4.6	25.6	107.6	84.7	192.3
	5.0% sulfur coal												
5	Limestone	5.7	10.7	22.6	41.1	-	7.2	5.9	5.6	40.3	139.1	107.6	246.7
44	Lime	6.3	3.9	21.3	35.8	-	7.1	5.4	4.8	33.7	118.3	92.3	210.6
	Venturi - Spray Tower[b]												
	Without forced oxidation (pond)												
7	Limestone	5.0	9.2	26.8	47.6	-	7.1	5.2	6.1	29.3	136.3	107.1	243.4
46	Lime	4.8	3.4	25.0	39.6	-	6.9	4.9	5.1	25.6	115.3	91.0	206.3
	With forced oxidation (landfill)												
18	Limestone	5.0	9.3	26.7	46.5	2.7	7.1	7.2	6.2	9.2	119.9	98.5	218.4
49	Lime	4.8	3.4	24.9	38.9	2.1	6.9	6.8	5.3	8.4	101.5	83.9	185.4
	With forced oxidation and MgO (landfill)[c]												
39	Limestone	5.1	9.3	27.0	46.7	6.2	7.1	7.3	6.5	9.0	124.2	101.9	226.1
52	Lime	4.8	3.5	25.2	38.3	6.6	6.9	6.9	5.5	8.3	106.0	89.4	193.5

a. 500-MW, without forced oxidation, pond disposal.
b. 500-MW, 3.5% sulfur coal.
c. Bleedstream forced oxidation.

TABLE 39. ANNUAL REVENUE REQUIREMENTS COMPARISON FOR LIMESTONE AND LIME PROCESSES

Case	Process	First-year annual revenue requirements, 1984 mills/kWh											Levelized annual revenue requirements, 1984 mills/kWh
		Absorbent	Additive	Labor	Steam	Water	Electricity	Diesel fuel	Maintenance[a]	Overheads	Capital charges	Total	
	Spray Tower[b]												
	0.7% sulfur coal												
2	Limestone	0.07	-	0.14	-	0.01	0.53	-	1.13	0.77	3.91	6.56	8.91
41	Lime	0.25	-	0.14	-	0.01	0.32	-	0.94	0.65	3.25	5.56	7.62
	2.0% sulfur coal												
3	Limestone	0.23	-	0.16	0.50	0.01	0.48	-	1.34	0.90	4.83	8.46	11.68
42	Lime	0.85	-	0.16	0.32	0.01	0.37	-	1.17	0.80	4.24	7.92	11.18
	3.5% sulfur coal												
4	Limestone	0.44	-	0.17	0.49	0.01	0.68	-	1.60	1.06	6.02	10.47	14.42
43	Lime	1.64	-	0.17	0.50	0.01	0.44	-	1.37	0.92	5.14	10.19	14.66
	5.0% sulfur coal												
5	Limestone	0.63	-	0.17	0.49	0.02	0.73	-	1.69	1.12	6.59	11.44	15.73
44	Lime	2.34	-	0.17	0.49	0.01	0.45	-	1.45	0.97	5.63	11.51	16.71
	Venturi - Spray Tower[c]												
	Without forced oxidation (pond)												
7	Limestone	0.41	-	0.18	0.45	0.01	0.91	-	1.75	1.16	6.50	11.37	15.68
46	Lime	1.64	-	0.18	0.45	0.01	0.62	-	1.48	0.99	5.51	10.88	15.63
	With forced oxidation (landfill)												
18	Limestone	0.41	-	0.47	0.45	0.01	0.98	0.04	1.70	1.30	5.84	11.20	15.96
49	Lime	1.64	-	0.44	0.45	0.01	0.67	0.04	1.44	1.12	4.96	10.78	15.92
	With forced oxidation and MgO (landfill)[d]												
39	Limestone	0.39	0.02	0.47	0.45	0.01	0.97	0.04	1.76	1.35	6.04	11.50	16.33
52	Lime	1.57	0.01	0.44	0.45	0.01	0.64	0.04	1.50	1.17	5.17	11.01	16.18

a. Includes analysis.
b. 500-MW, without forced oxidation, pond disposal.
c. 500-MW, 3.5% sulfur coal.
d. Bleedstream forced oxidation.

comparisons based on coal sulfur content are for 500-MW spray tower processes without forced oxidation or additives. The comparisons based on process type (without forced oxidation, and with forced oxidation with and without magnesium oxide) are 500-MW, 3.5% sulfur coal, venturi - spray tower processes.

The processes using lime maintain a substantial advantage in capital investment for all conditions. The advantage is relatively constant with increasing coal sulfur content (the lime process is 17% lower in capital investment for the 0.7% sulfur coal and nearly 15% lower for the 5% sulfur coal compared with the corresponding limestone process). In terms of process type, the capital investment advantage of the lime processes also remains essentially constant at 14% to 15% lower than the corresponding limestone processes.

The lower capital investments for lime processes are primarily the result of lower costs in the feed preparation, SO_2 absorption, and disposal areas. These accrue because of the absence of grinding equipment, improved SO_2 absorption, and high absorbent utilization. Additional, although smaller, savings are achieved in the gas handling, solids separation, and reheat areas.

In annual revenue requirements, coal sulfur content has an important effect on the cost relationships of the limestone and lime processes. Electricity costs and capital charges are lower for the lime process but increasing absorbent costs as the coal sulfur content increases result in higher annual revenue requirements for the lime process for the 5.0% sulfur coal case. Comparing the spray tower processes, for the 0.7% sulfur case the lime process has a 14% cost advantage in first-year annual revenue requirements over the limestone process. This advantage decreases with increasing sulfur content until at approximately 4.5% sulfur coal, limestone becomes less expensive. For the 5.0% sulfur case limestone is about 1% lower in first-year annual revenue requirements. In terms of levelized annual revenue requirements, where the effects of inflation and the cost of money on operating and maintenance costs are considered, the costs of lime and limestone processes are approximately equivalent at a coal sulfur level of about 3.0% with the lime process being less expensive at lower sulfur levels (more than 14% lower for the 0.7% sulfur coal case).

The annual revenue requirements relationships of limestone and lime processes are also dependent on the type of absorber and the use of forced oxidation and additives. The use of high-pressure-drop absorbers, such as the venturi - spray tower, slightly increases the advantage of lime processes. For the 3.5% sulfur coal case forced oxidation reduces the first-year annual revenue requirements of both limestone and lime processes slightly but the effect is somewhat greater for the limestone process. Forced oxidation and the use of magnesium oxide together result in a slight increase in first-year annual revenue requirements for both limestone and lime processes. For both processes the increase is essentially the same. In terms of levelized annual revenue requirements the costs for limestone and lime processes are equivalent (less than a 1% difference) in all three of the cases. The levelized costs increase slightly (about 2%) in going from the process without forced oxidation to the forced-oxidation process and increase again (about 2%) in going to magnesium oxide addition.

Cost Sensitivities

Most FGD processes differ from one another to some degree in cost distribution among the various components of annual revenue requirements. The sensitivity of the total annual revenue requirements to changes of their component costs depends on the percentage of the total annual revenue requirements represented by the component cost. Variations in the relative costs of different processes may occur if the sensitivities of the processes to particular component cost changes differ. The effect is greatest between highly dissimilar processes in which the component costs differ greatly or are unique to one process.

Table 40 shows the distribution of cost elements in annual revenue requirements for comparable limestone processes and a lime process. The largest components of the annual revenue requirements of limestone processes are capital charges, followed by maintenance, overheads, electricity, steam, limestone, and labor. There is relatively little difference in the cost distribution of the annual revenue requirements for the various limestone processes. The greatest differences in cost distribution are between the processes without forced oxidation and processes with forced oxidation. The processes without forced oxidation (all with pond disposal) have relatively lower labor, electricity, and overhead costs and higher capital charges. Costs related to forced oxidation and the use of additives are too small a percentage of total annual revenue requirements to affect the relative economics of the processes even if radical variations in them were to occur. In the case of lime processes, however, the absorbent cost is a major factor--exceeding all other cost elements except capital charges--and the capital charges constitute a smaller portion of the total than they do in limestone processes. In other component costs the lime processes are similar to limestone processes.

The sensitivity of limestone process annual revenue requirements to changes in important cost elements is shown in Table 41. Changes in capital charges are shown as changes in the inflation rate and discount rate (the levelizing factor) rather than as changes in capital investment, which has a more obvious and direct effect on capital charges. Changes in the discount rate also affect capital investment through its effect on interest during construction. Changes in the operating and maintenance portion of annual revenue requirements also have a slight effect on capital investment through working capital. Therefore, the effect on capital investment is also shown in Table 41.

Changes in Inflation/Discount Rate--

Changes in the annual inflation and discount rates from 4% inflation and a 7% discount rate to 8% inflation and a 13% discount rate (compared with the 6% inflation and a 10% discount rate used in this study) have a moderate

TABLE 41. COST SENSITIVITIES

	Process[a]		
	Case 4	Case 15	Case 27
Base Case[b]			
Capital investment, $/kW	225.1	206.1	185.5
Annual revenue requirements, mills/kWh			
First-year	10.47	10.58	9.38
Levelized	14.42	15.08	13.31
4% Inflation, 7% Discount			
Capital investment, $/kW	216.7	198.4	178.6
Annual revenue requirements, mills/kWh			
First-year	8.75	9.00	7.97
Levelized	11.44	12.07	10.64
8% Inflation, 13% Discount			
Capital investment, $/kW	233.7	214.0	192.7
Annual revenue requirements, mills/kWh			
First-year	11.63	11.64	10.34
Levelized	16.70	17.42	15.38
Maintenance x 1.5			
Capital investment, $/kW	226.0	207.0	186.3
Annual revenue requirements, mills/kWh			
First-year	11.74	11.85	10.53
Levelized	16.79	17.45	15.45
Maintenance x 0.5			
Capital investment, $/kW	224.2	205.3	184.8
Annual revenue requirements, mills/kWh			
First-year	9.20	9.31	8.24
Levelized	12.04	12.70	11.17
Labor x 1.5			
Capital investment, $/kW	225.2	206.4	185.8
Annual revenue requirements, mills/kWh			
First-year	10.64	10.99	9.75
Levelized	14.73	15.84	13.99
Labor x 0.5			
Capital investment, $/kW	225.0	205.9	185.3
Annual revenue requirements, mills/kWh			
First-year	10.30	10.17	9.02
Levelized	14.10	14.32	12.62

a. All are 500 MW, 3.5% sulfur coal, spray tower processes.
Case 4: limestone, no forced oxidation, pond disposal.
Case 15: limestone, forced oxidation, landfill disposal.
Case 27: limestone, forced oxidation, adipic acid additive, landfill disposal.

b. 6% inflation, 10% discount rate.

effect on capital investment and an appreciable effect on annual revenue requirements. The capital investment for the limestone forced-oxidation process increases about 8%, the first-year annual revenue requirements increase about 30%, and the levelized annual revenue requirements increase about 45% between the lower and higher rates. The effects are shown in Figure 20.

TABLE 40. COST DISTRIBUTION AS A PERCENTAGE

OF TOTAL FIRST-YEAR ANNUAL REVENUE REQUIREMENTS

	Process[a]			
	Case 4	Case 15	Case 27	Case 47
Absorbent	4	4	4	16
Additive	-	-	<1	-
Labor	2	4	4	4
Steam	5	5	4	5
Process water	<1	<1	<1	<1
Electricity	6	7	7	5
Diesel fuel	-	<1	<1	<1
Maintenance	15	15	15	13
Analysis	<1	<1	<1	<1
Overheads	10	12	12	10
Capital charges	58	52	53	46

a. All are 500 MW, 3.5% sulfur coal, spray tower processes.
Case 4: limestone, no forced oxidation, pond disposal.
Case 15: limestone, forced oxidation, landfill disposal.
Case 27: limestone, forced oxidation, adipic acid additive, landfill disposal.
Case 47: lime, forced oxidation, landfill disposal.

Changes in Labor and Maintenance Costs--

The sensitivity of process costs to changes in labor and maintenance costs is shown in Figure 21, also for the limestone forced-oxidation process. Changes in labor costs have very little effect because of the small fraction of annual revenue requirements that constitute labor costs. The effect of changes in maintenance costs is greater, reflecting the 15% fraction of annual revenue requirements composed of maintenance costs.

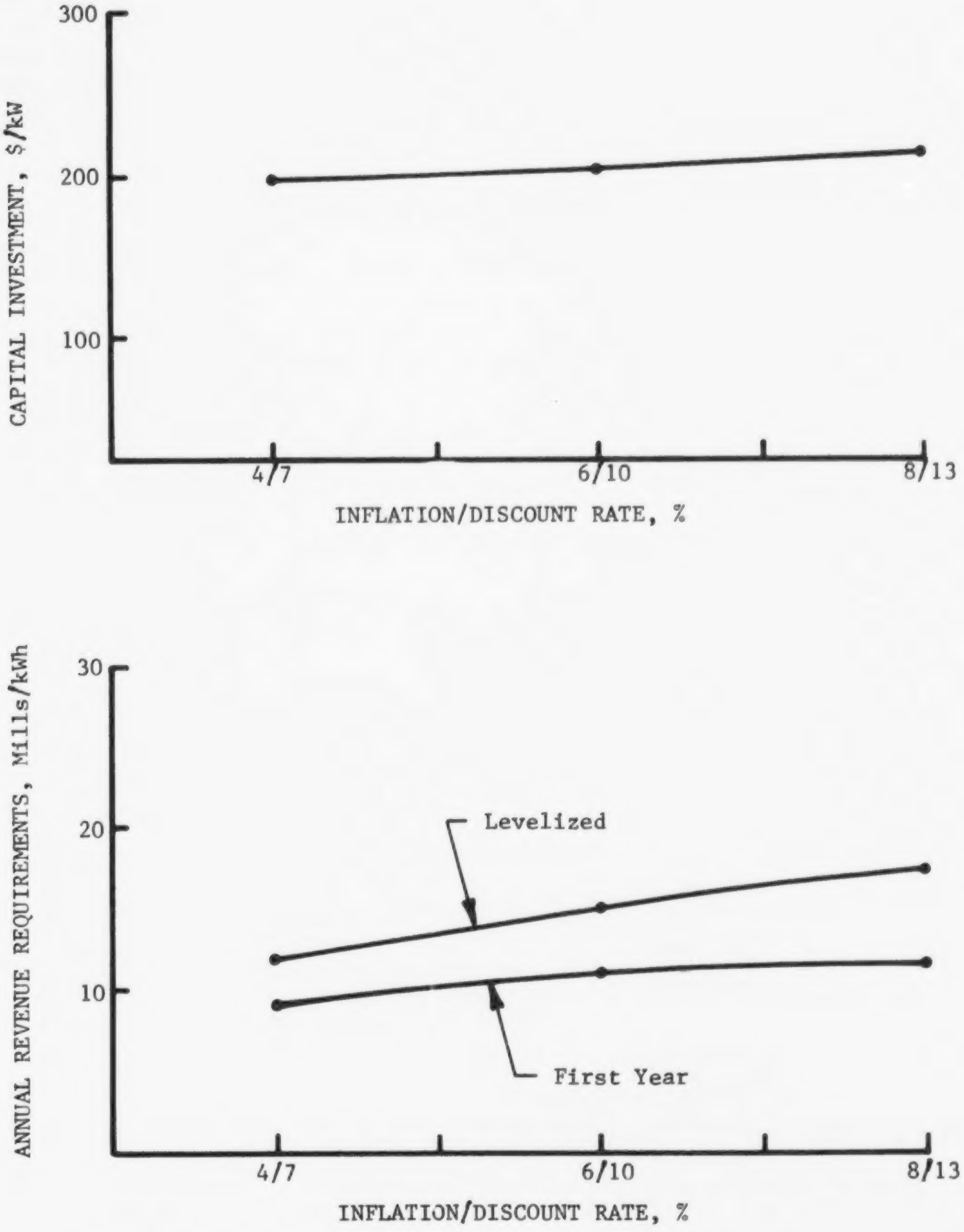


Figure 20. Sensitivity of limestone forced-oxidation process costs to changes in inflation and discount rate.

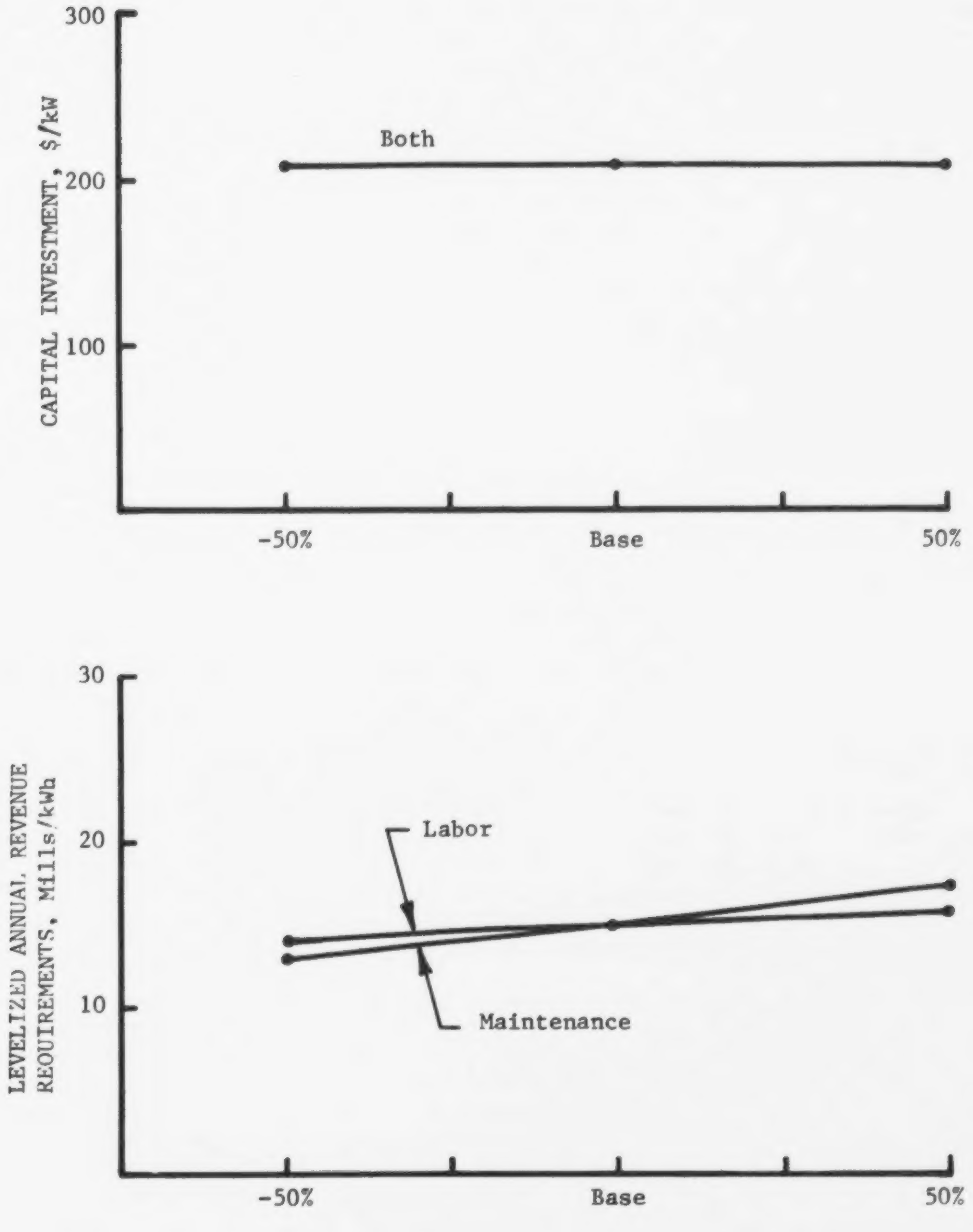


Figure 21. Sensitivity of limestone forced-oxidation process costs to changes in labor and maintenance costs.

An additional facet of maintenance costs lies in their effect on the relative economics of processes that differ in the type of absorber used. Maintenance costs in this study are quantified as a function of capital investment, which reflects the quantity and, in general, the type of equipment. Detailed operating data that would permit a more direct definition of maintenance costs are not presently available in useful form. Therefore, the same maintenance cost factor is used for all three types of absorbers evaluated. As was discussed, processes with a TCA have lower capital investments and annual revenue requirements than comparable processes with a spray tower. If maintenance costs are proven to differ significantly between TCA's and spray towers, the relationship of the annual revenue requirements would change. Figure 22 illustrates the effect of change in maintenance costs on the relationships of annual revenue requirements for TCA and spray tower forced-oxidation processes. A change of about 15% in maintenance costs (about $600,000/yr, 0.2 mills/kWh) from those used in this study eliminates the difference in annual revenue requirements between the two processes.

Changes in Lime Costs--

In contrast to the limestone processes, absorbent cost is an appreciable element of the annual revenue requirements for the lime process. The higher sensitivity of the lime process to absorbent costs is illustrated in Figure 23 for a lime forced-oxidation process. As would be expected, the effect on capital investment is relatively insignificant. The effect on both first-year and levelized annual revenue requirements is, however, significant. A 50% increase in the delivered lime cost results in an 8% increase in first-year annual revenue requirements and a 21% increase in levelized annual revenue requirements.

Use of Dibasic Acid Instead of Adipic Acid--

Questions have been raised about the effects of the high cost and degradation rate of adipic acid on the process economics for full-scale utility FGD applications. There have been suggestions about using dibasic acid (DBA), which is a byproduct of adipic acid manufacture, as an alternative to adipic acid because of its lower unit cost ($640/ton, dry basis versus $1,500/ton, dry basis for adipic acid).

Table 42 compares the process economics of limestone processes, one enhanced with adipic acid and the other enhanced with DBA instead of adipic acid. Both are forced-oxidation processes for a new 500-MW unit burning 3.5% sulfur coal. It is assumed that the use of DBA results in improvements in scrubber operation identical to improvements resulting from the use of adipic acid. The DBA, however, appears to have a higher degradation rate (approximately one-third higher) than adipic acid and thus, the DBA addition rate is higher in order to maintain the same additive concentration in the scrubbing slurry (174).

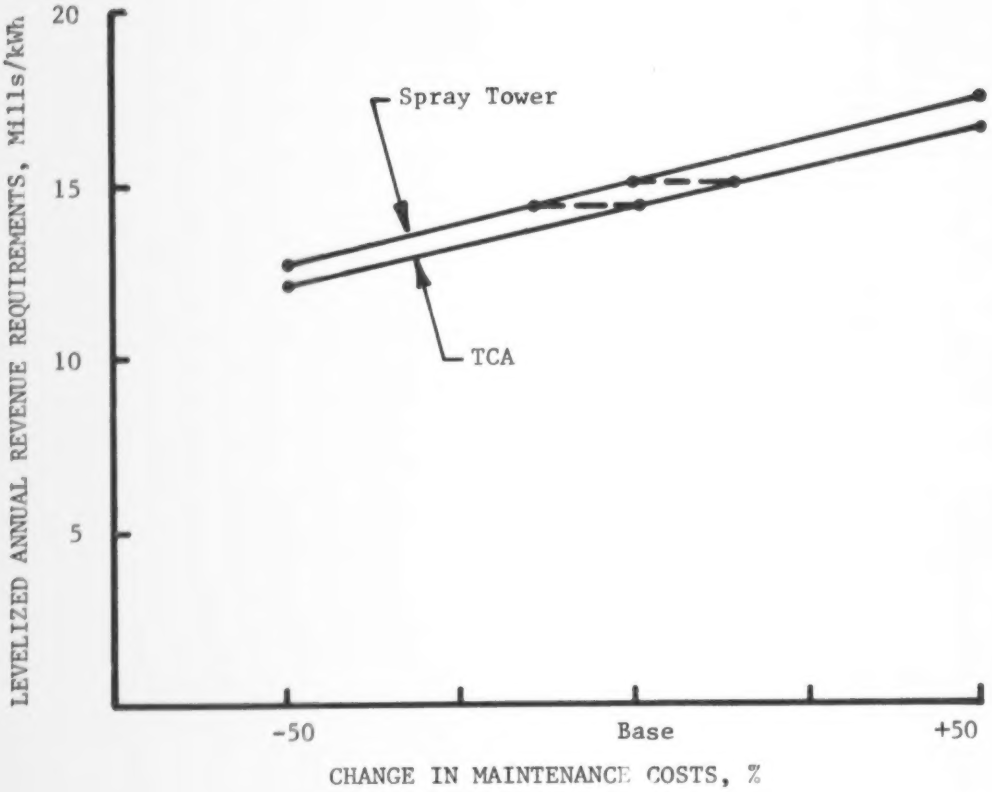


Figure 22. Annual revenue requirement variation with maintenance costs for limestone TCA and spray tower landfill processes.

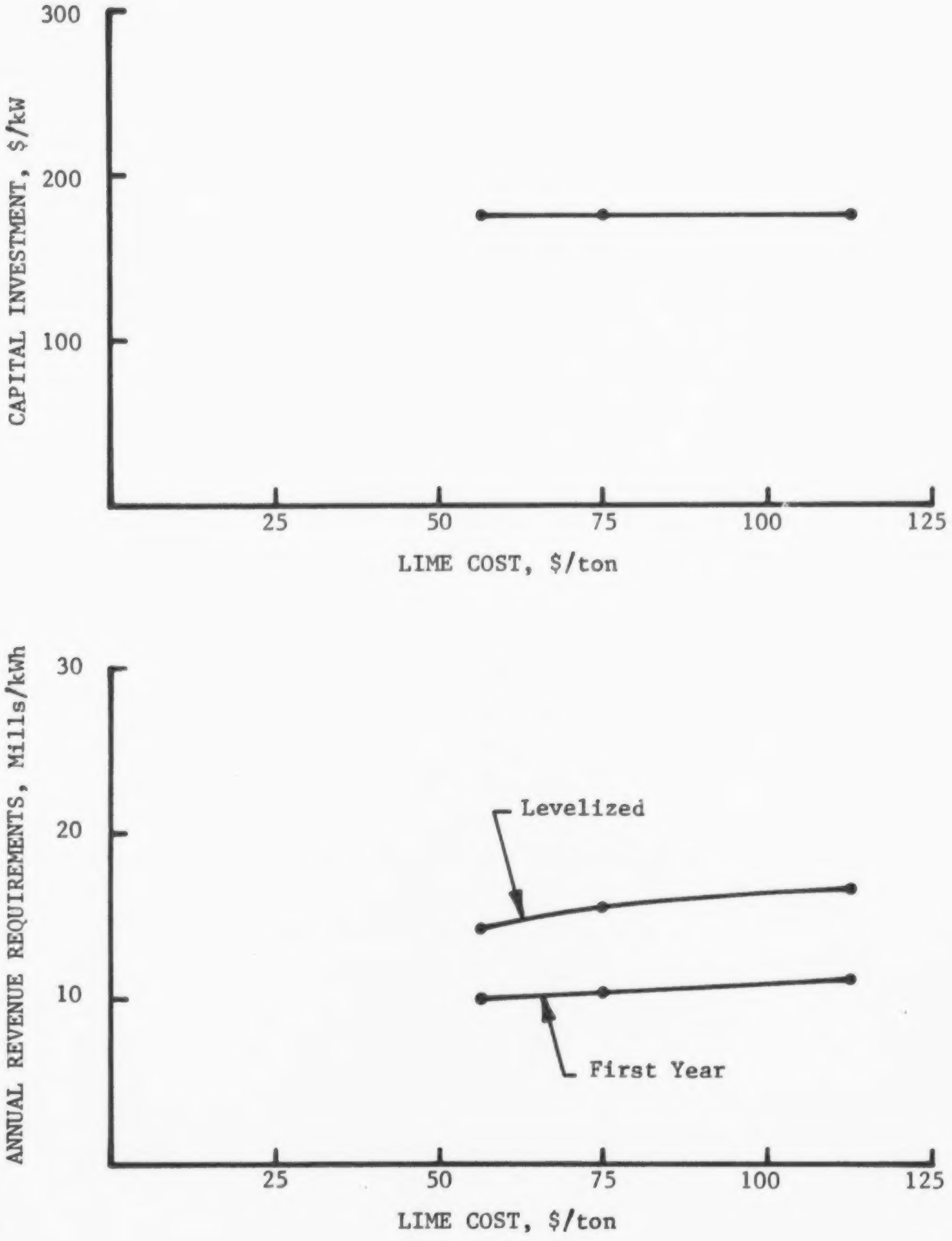


Figure 23. Sensitivity of lime process costs to the cost of lime.

TABLE 42. COST COMPARISON OF ADIPIC-ACID- AND DBA-ENHANCED LIMESTONE PROCESSES

	Capital investment, 1982$		Annual revenue requirements, 1984$			
			First-year		Levelized	
Process[a]	M$	$/kW	M$	Mills/kWh	M$	Mills/kWh
Limestone with adipic acid	92.8	186	25.81	9.38	36.59	13.31
Limestone with dibasic acid	92.8	186	25.73	9.35	36.44	13.25

a. 500 MW, 3.5% sulfur coal, spray tower, forced oxidation, landfill.

As can be seen, the capital investments for the two processes are essentially the same and the annual revenue requirements are only marginally lower (less than 0.5%) for the DBA case. Although the cost for DBA is much lower than for adipic acid, the low consumption for both additives and the higher degradation rate for DBA result in only a slight, probably negligible, cost advantage for DBA. (However, more test work with DBA is needed to confirm these design assumptions, particularly since DBA enhancement has never been tested with forced-oxidation.)

Changes in Adipic Acid Cost--

For this study a delivered cost for adipic acid of $1,500/ton was used. Since adipic acid is derived from petroleum and thus may be subject to large increases in cost, a sensitivity analysis was performed to demonstrate the effect of 20% changes in the delivered cost of adipic acid on the economics of an adipic-acid-enhanced limestone process. The results are shown in Table 43 and Figure 24. The capital investment and the levelized annual revenue requirements are essentially constant, increasing only 0.05% and 0.15%, respectively, for a 20% increase in the delivered cost of adipic acid. Thus, although adipic acid is expensive, the consumption is low and the cost is negligible when compared with the other costs.

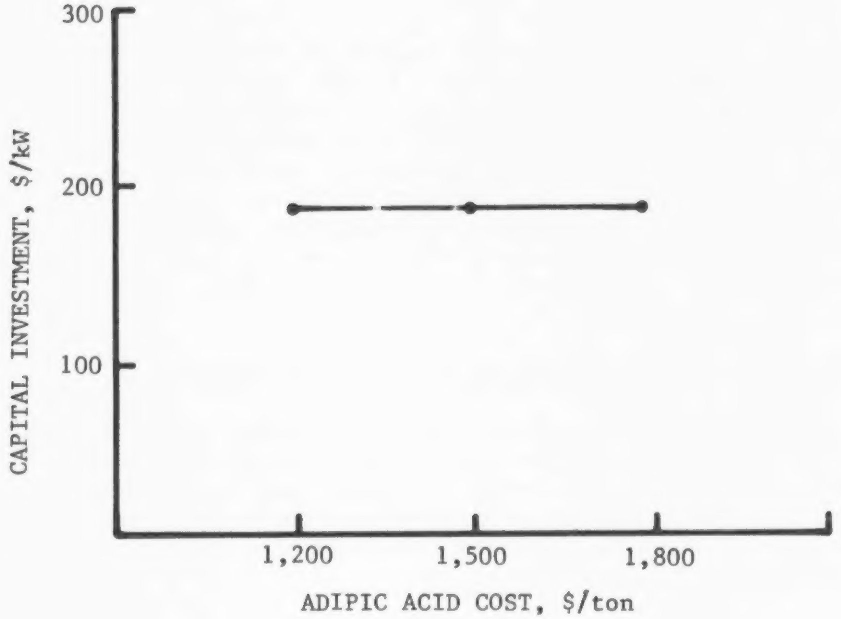
Changes in Adipic Acid Degradation Rate--

Adipic acid tends to degrade in the absorber into a mixture of primarily valeric and glutaric acids but with lesser amounts of butyric and succinic acids, carbon dioxide, methane, and butane. Some adipic acid is also lost with the waste. In order to simplify the model, adipic acid losses due to degradation and losses in the waste are combined as a single factor. The makeup adipic acid rate to replace adipic acid lost by degradation is assumed to be three times the rate of adipic acid loss in the waste. Therefore, the adipic-acid-enhanced limestone process as defined in this study is said to have a degradation factor of 3.

TABLE 43. SENSITIVITY OF LIMESTONE PROCESS COSTS TO ADIPIC ACID COST AND CONSUMPTION

	Limestone process[a]						
	Adipic acid cost, $/ton			Adipic acid degradation factor			Worst case: Cost of $1,800/ton and degradation factor of 9
	1,200	1,500[b]	1,800	3[b]	6	9	
Capital investment, $/kW	185.5	185.5	185.6	185.5	185.6	185.7	185.8
Annual revenue requirements, mills/kWh							
First-year	9.37	9.38	9.40	9.38	9.46	9.54	9.59
Levelized	13.28	13.31	13.33	13.31	13.45	13.60	13.69

a. 500-MW, 3.5% sulfur, limestone spray tower, forced-oxidation, landfill processes.
b. Base-case conditions.



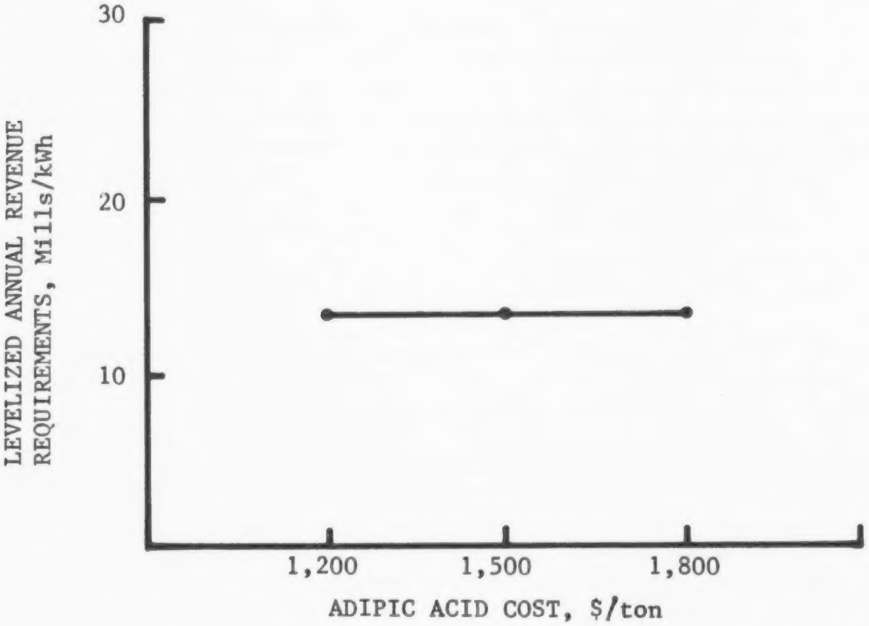


Figure 24. Sensitivity of limestone adipic acid forced-oxidation process costs to the cost of adipic acid.

However, adipic acid use in full-scale limestone FGD systems is relatively new and sufficient data are not yet available to predict adipic acid losses accurately for specific cases. Therefore, a sensitivity analysis was performed to illustrate the change in process economics as the adipic acid degradation factor is increased threefold (from an initial value of 3 to a worst case value of 9). The results are shown in Table 43 and Figure 25. Tripling the degradation factor increases the capital investment about 0.1% and increases the levelized annual revenue requirements about 2.2%. Thus, the consumption even at high degradation rates results in only a small increase in process costs.

Combined Effect of Adipic Acid Cost and Degradation Factor--

A worst case example with both the highest unit cost and the highest degradation factor for adipic acid is also shown in Table 43. Even in this worst case the capital investment increases only 0.2% and the levelized annual revenue requirements increase only 2.9%. Thus, within reasonable limits, the delivered cost for adipic acid and the degradation factor have little effect on the process economics.

COMPARISON OF PREVIOUS AND CURRENT TVA LIMESTONE FGD ECONOMICS

Since the limestone FGD process has been evaluated in several earlier TVA economic evaluations, it is possible to compare the results of previous evaluations with those in this study. This is instructive for several reasons. The differences in economics reflect on one hand technological changes in limestone FGD that have occurred over several years as well as refinements of power plant fuel and operating conditions upon which the FGD designs are based. The differences also reflect economic changes (inflation) and further refinements of the economic premises used to determine costs. Further, many of the economic differences are independent of a particular FGD process and represent general changes in FGD technological and economic concepts. The overall changes are typified by the changes in capital investment over the period during which the TVA economic evaluations have been made, as shown below.

The inflationary effects are obvious but the technological effects are perhaps more impressive, particularly since the 1969 (175) and 1975 (176) costs include fly ash removal as well as SO_2 removal.

The overall changes can be shown in more detail by a point-by-point comparison of the results of this study with the two most recent preceding studies. In the following comparison the standard limestone process as evaluated in the most recent preceding TVA economic evaluations (1,2), which were also determined using the Shawnee computer program, is compared with the standard limestone process of this study by a stepwise conversion from the old design and economic premises to the new design and economic premises used in

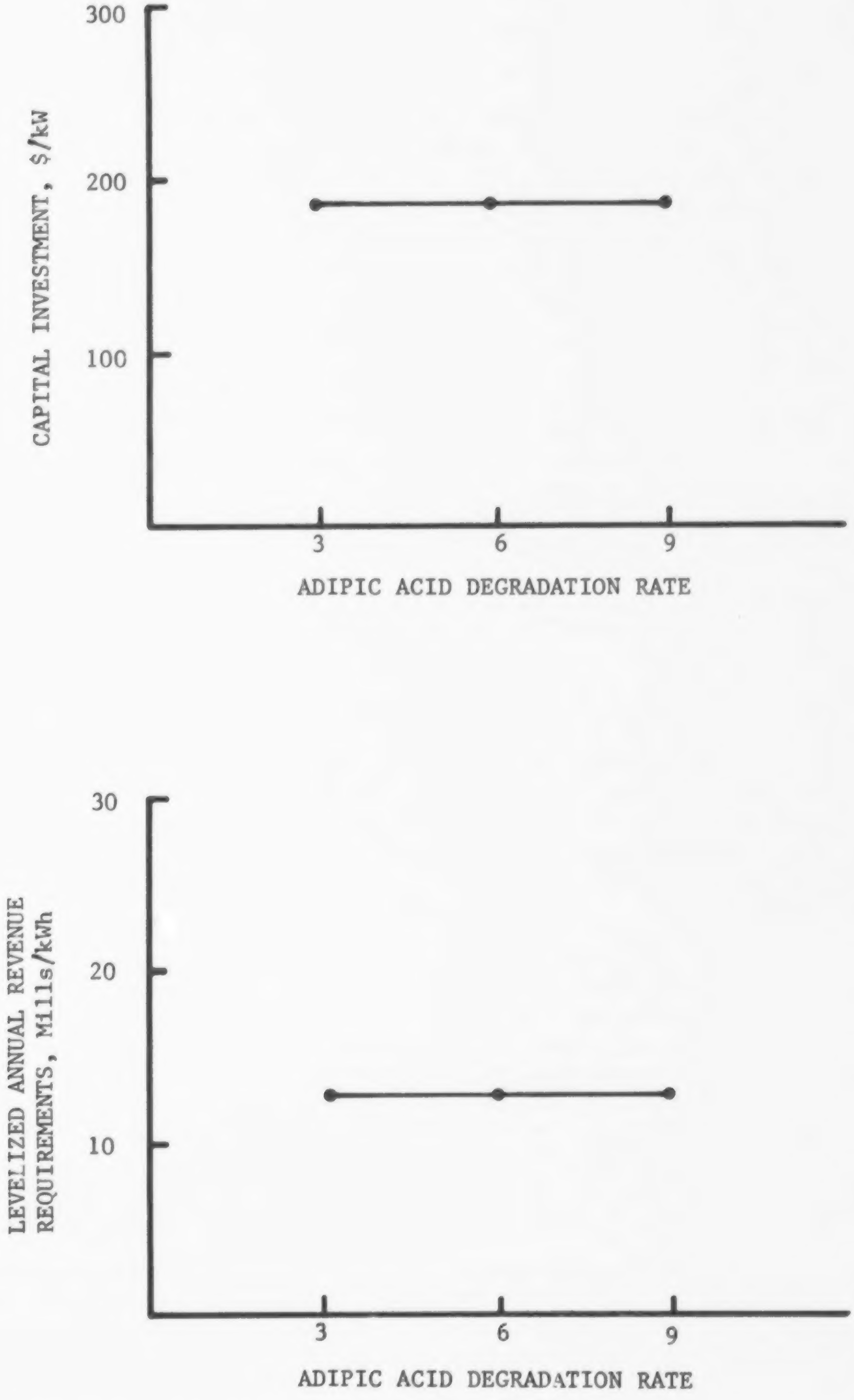
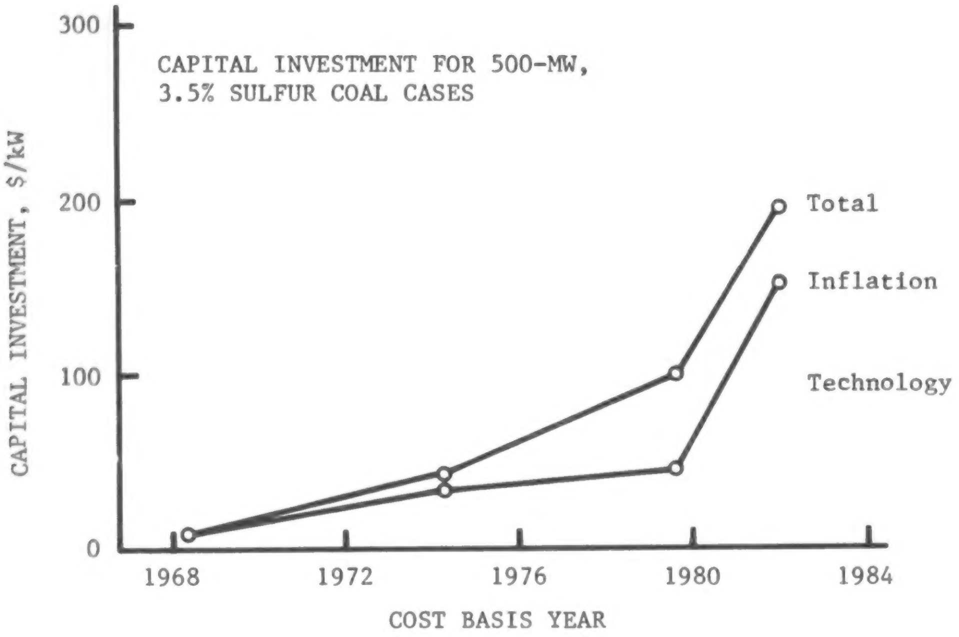


Figure 25. Sensitivity of limestone adipic acid forced-oxidation process costs to the adipic acid degradation rate.



this study. The comparison evolves the standard process from the preceding evaluation to the standard process from this study, a spray tower process with in-loop forced oxidation and landfill disposal. The major differences between the old and new premises are discussed below.

Design Premises

The base-case power unit remains essentially the same; i.e., north-central location, a new 500-MW, pulverized-coal-fired, dry-bottom boiler. The heat rate is increased from 9,000 to 9,500 and the excess air is increased from 33% to 39%, however. The sulfur content of the coal remains at 3.5% but the heating value is increased from 10,500 to 11,700 Btu/lb. The operating schedule is also changed to 5,500 hr/yr for 30 years or 165,000 total hours over the life of the plant. In the old design premises a declining operating schedule beginning with 7,000 hr/yr and a total operating life of 127,500 hours over the 30-yr life of the plant was used. A constant annual operating time is used to facilitate levelizing of annual revenue requirements.

Major changes in the FGD design premises reflect current regulations and changes to improve process reliabilities. Required SO_2 removal efficiency is now based on the 1979 NSPS. For the base-case coal these require an 89% removal efficiency instead of the 79% needed to meet the 1971 NSPS used in the old premises. In keeping with current design trends a spare absorber train and provisions for emergency bypass are included. The old premises contained no spare absorber or bypass provisions. In addition, ID booster fans, instead of FD booster fans, are used in the new premises. Both pond and landfill waste disposal methods are revised to reflect more recent environmental concerns. These are primarily based on RCRA Subtitle D (nonhazardous waste) guidelines and include provisions for such factors as seepage and runoff control, security, monitoring, and reclamation.

The current standard limestone process differs from the old process used in the preceding studies in several respects. A spray tower instead of a TCA, forced oxidation to gypsum, and landfill waste disposal instead of pond disposal are now used in the basic system. The use of a spray tower results in a lower gas velocity of 10 ft/sec instead of the 12.5 ft/sec used in the old process with a TCA.

The new limestone scrubbing process represents several industry trends in limestone scrubbing that have become evident in recent years. The use of a spray tower instead of more complicated TCA or venturi - spray absorbers has become common (90). The simplier spray tower is expected to provide greater reliability and require less maintenance although these have not been quantified in practice. The problem of waste disposal has also been addressed by producing a more tractable waste, by oxidation to gypsum.

Economic Premises

Numerous changes were also made in the economic premises. Specific provisions for sales tax, freight, and overtime for construction delays are included. The method of calculating indirect capital investment is simplified and modified to more accurately reflect the engineering and construction costs of the processes. Contingencies and allowances for modification after startup are also defined as process-specific variables reflecting the degree of development of the technology. Provision for recognition of anticipated royalties is also made. Land prices and interest during construction are increased.

First-year annual revenue requirements are now calculated using levelized capital charges (30-year life, sinking fund depreciation, 6% per year inflation and 10% per year cost of money, discounted to the first year) instead of the average capital charges used in the old premises. In addition, levelized annual revenue requirements are also calculated to represent inflated and discounted costs over the life of the system.

The base years for capital investment and first-year revenue requirements are also advanced to 1982 and 1984 respectively. A project construction period from 1981 to 1983 is now assumed, with plant startup in early 1984.

Cost Comparison

The key old and new design and economic premises for evaluation of the limestone process are shown in Table 44. A stepwise cost transition from

TABLE 44. COMPARISON OF OLD AND NEW PREMISE CONDITIONS FOR THE LIMESTONE PROCESS

	Old premises	New premises
Design Premises		
Coal, Btu/lb	10,500	11,700
Excess air, %	33	39
Heat rate, Btu/kWh	9,000	9,500
Operating profile		
First year, hr/yr	7,000	5,500
Lifetime, hr (30 years)	127,500	165,000
FGD		
SO_x removal, %	1971 NSPS	1979 NSPS
Emergency bypass, %	0	50
Spare units	0	1
Booster fan	FD	ID
Limestone process		
Absorber	Mobile bed	Spray tower
L/G, gal/kaft^3	50	106
Gas velocity, ft/sec	12.5	10.0
ΔP, in. H_2O	8	1.4
Forced oxidation	No	Yes
Waste disposal	Pond	Landfill
Economic Premises		
Cost index year		
Capital investment	1979	1982
Annual revenue requirements	1980	1984
Indirect capital costs		Revised
Land, $/acre	3,500	5,000
Interest during construction, %	12	15.6
Limestone process contingency, %	20	10
Waste disposal contingency, %	20	20
Waste disposal allowance for startup, %	8	0
Capital charges	Average annual	Levelized
Depreciation	Straight line	Sinking fund

the old premises and technology to those for the new limestone scrubbing evaluation is shown in Table 45 and illustrated in Figure 26. Overall, the cumulative changes more than double the capital investment and first-year revenue requirements. The largest cost increase is a result of advancing the cost index year from 1979 to 1982 for capital investment and from 1980 to 1984 for first-year revenue requirements.

The capital investment increases resulting from the new economic premises are related to higher indirect capital investment costs, particularly in interest during construction, contractor expense, and working capital. The increase in first-year revenue requirements stems largely from capital charges based on the capital investment.

New power plant coal and air rates, the operating profile, and the 1979 NSPS all produce similar increases in capital investment. In these cases the main factors are increased flue gas volume, increased lifetime waste disposal requirements, and the more stringent SO_2 removal conditions. The effect on annual revenue requirements is similar except, of course, that the reduction in yearly operating hours results in a reduction in annual costs ($/yr) and an increase in unit costs (mills/kWh). Addition of reliability factors (a spare absorber train, emergency bypass, and a spare ball mill) also cause appreciable increases in both capital investment and first-year revenue requirements. The use of a spray tower instead of a TCA increases costs primarily because of the lower flue gas velocity and higher slurry recirculation rate, which requires larger ducting and pumping requirements. Substitution of landfill for ponding substantially reduces capital investment by eliminating pond construction costs. The resulting reduction in capital charges helps to counteract the increased waste disposal costs in first-year annual revenue requirements.

Overall, economics in the form of inflation and higher interest rates have the largest effect in comparison of the limestone process using the old and new premises and technology. Technical changes related to improvements in reliability, such as bypass and redundancy provisions, also have a large effect. The higher SO_x removal efficiency has a much less effect than the economic and technical changes.

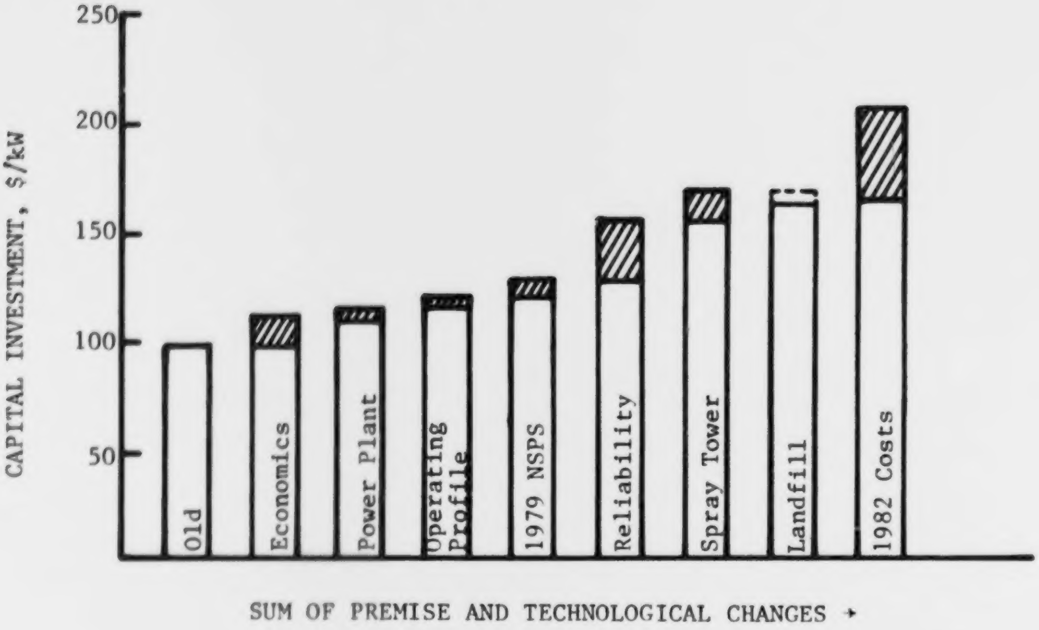
TABLE 45. STEPWISE CONVERSION OF LIMESTONE PROCESS COSTS

FROM OLD TO NEW PREMISES AND TECHNOLOGY

	Capital investment				First-year revenue requirements			
Condition	k$	$/kW	% change	% total change	k$	Mills/kWh	% change	% total change
Old premises and technology	48,700	98	-	-	14,100	4.0[a]	-	-
Above with new economic premises and pond	55,100	110	13	13	16,200	4.6[a]	15	15
Above with new power plant design premises	57,100	114	4	17	17,000	4.9[a]	5	21
Above with new operating profile	59,800	120	5	23	16,500	6.0[b]	-3	17
Above with 1979 NSPS	63,600	127	6	29	17,200	6.3[b]	4	22
Above with reliability factors (spares and bypass)	77,100	154	21	58	20,100	7.3[b]	17	43
Above with spray tower	83,300	167	8	71	21,500	7.8[b]	7	52
Above with landfill	81,400	163	-2	67	23,000	8.4[b]	7	63
Above with 1982, 1984 costs	103,100	206	27	112	29,100	10.6[b]	27	106

a. Based on 7,000 hr/yr.

b. Based on 5,500 hr/yr.



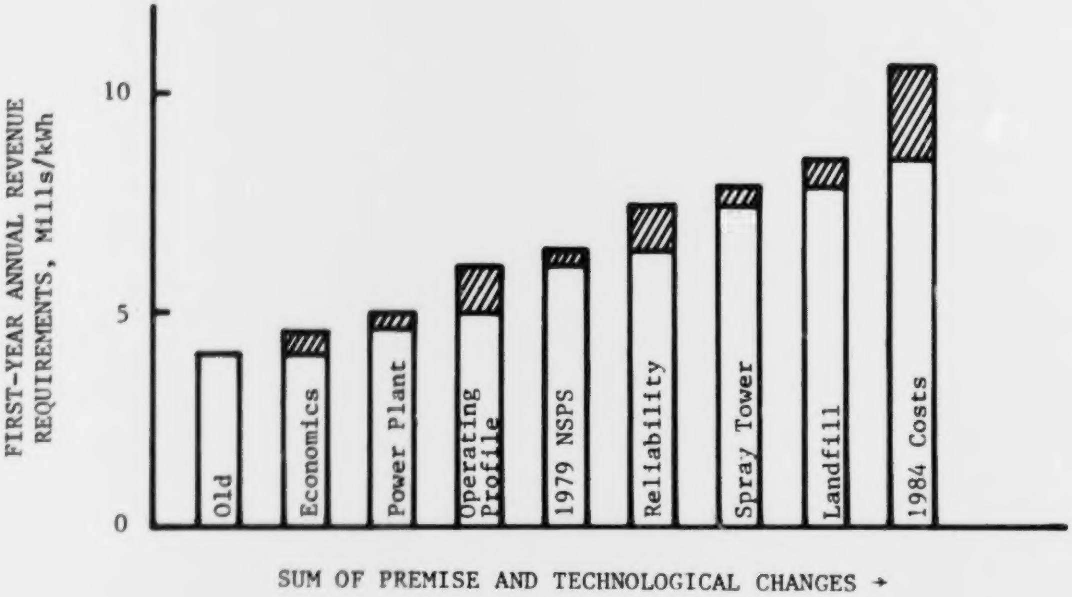


Figure 26. Stepwise conversion of limestone process costs from old to new premises and technology.

CONCLUSIONS

Processes with forced oxidation and landfill disposal of solid wastes have lower capital investments than processes without forced oxidation and pond disposal because the aggregate costs of forced oxidation and dewatering equipment and landfill construction are small, compared with the cost of pond construction. First-year annual revenue requirements are slightly higher for forced-oxidation processes under most conditions evaluated. Processes without forced oxidation have slightly lower levelized annual revenue requirements in all cases.

The use of additives reduces both the capital investment and annual revenue requirements of limestone processes. For processes with forced oxidation and landfill disposal, the effects of magnesium oxide are marginal--a reduction of 2.5% or less--but with adipic acid more substantial reductions of 10% to 13% in both capital investment and annual revenue requirements are obtained. The cost reductions are a result of more efficient SO_2 removal and absorbent utilization. The costs related to the use of additives are a very small part of the total costs and both capital investment and annual revenue requirements are very insensitive to changes in the cost and consumption rate of the additives. For this reason, the use of lower cost substitutes such as DBA has almost no effect on overall costs. For processes that use pond disposal, the use of additives has the same effect on capital investment as it does for processes using landfill disposal but the cost reduction in annual revenue requirements is reduced about one-half because of the additional additive loss in the low-solids waste.

Processes using TCA's are about 5% less expensive than processes using spray towers and 12% less expensive than processes using venturi - spray towers because of the smaller size and lower L/G ratio of the high-pressure-drop TCA, assuming that maintenance costs are the same percentage of equipment costs for all the processes. Changes of 15% in maintenance costs for either process make the annual revenue requirements of processes using TCA's and spray towers equivalent.

Economically, the optimum absorber design favors higher limestone stoichiometries and lower L/G ratios. For the same SO_2 removal efficiency, a 1.4 stoichiometry and a low L/G are 14% to 18% lower in capital investment and annual revenue requirements than a design with a 1.1 stoichiometry and a high L/G ratio.

Partial scrubbing at 90% removal efficiency with bypass of some of the flue gas is more economical at coal sulfur contents below about 2% because the bypassed flue gas reduces reheat costs.

Lime processes have lower capital investments than limestone processes in all of the cases evaluated. They also have lower annual revenue requirements at lower coal sulfur contents. Lime processes are more sensitive to absorbent costs, however, and also to levelizing, since operating and maintenance costs constitute a larger percentage of the total annual revenue requirements. As a result, lime process annual revenue requirements increase more rapidly with coal sulfur content than those of limestone processes, particularly if levelized annual revenue requirements are compared. At a coal sulfur content of about 4.5% the first-year annual revenue requirements of lime and limestone processes are equal; at a coal sulfur content of about 3.0% the levelized annual revenue requirements are equal. As the coal sulfur content further increases, limestone processes become increasingly more economical. Except for minor differences these relationships are true for variations in processes such as the use of other absorbers, forced oxidation, and landfill.

The largest components of annual revenue requirements for limestone processes, regardless of type, are, in order of decreasing importance, capital charges, maintenance, overheads, electricity, steam, limestone, and labor. Capital charges constitute over one-half and overheads one-tenth of the total so the annual revenue requirements are quite sensitive to changes in capital investment, inflation, and the discount rate. Except for a moderate sensitivity to maintenance costs, the limestone processes are relatively insensitive to changes in the other cost elements. The largest components of annual revenue requirements for lime processes are, in order of decreasing importance, capital charges, absorbent, maintenance, overheads, electricity, steam. and labor. The annual revenue requirements of lime processes are also most sensitive to changes in capital investment, inflation, and the discount rate, but slightly less so than those of limestone processes because capital charges constitute slightly less than one-half of the annual revenue requirements. It is, however, moderately sensitive to absorbent cost as well as maintenance costs.

The costs of the base-case limestone process used in this study are approximately double those of the base-case limestone process used in similar previous studies. The differences are the result of updated design and economic premises used in this study. The largest cost increase, about 25% of the total, is the result of advancing the cost index year from 1979 to 1982 for capital investment and 1980 to 1984 for annual revenue requirements. Incorporation of spray towers, spare absorbers, and emergency bypass produced the second largest increase, also about 25% of the total. General revisions of the economic premises account for about 15% of the increase. The remaining premise changes (incorporation of forced oxidation, substitution of landfill for pond disposal, and substitution of 1979 NSPS for 1971 NSPS) have a minor impact and, individually, do not account for more than a 5% increase in costs.

REFERENCES

1. S. V. Tomlinson, F. M. Kennedy, F. A. Sudhoff, and R. L. Torstrick, 1979, Definitive SO_x Control Process Evaluations: Limestone, Double Alkali, and Citrate FGD Processes, EPA-600/7-79-177, U.S. Environmental Protection Agency, Washington, D.C.

2. K. D. Anderson, J. W. Barrier, W. E. O'Brien, and S. V. Tomlinson, 1980, Definitive SO_x Control Process Evaluations: Limestone, Lime, and Magnesia Processes, EPA-600/7-80-001, U.S. Environmental Protection Agency, Washington, D.C.

3. Journal of the Air Pollution Control Association, 1977, A History of Flue Gas Desulfurization Systems Since 1850, Vol. 27, No. 10, pp. 943-961

4. H. F. Johnstone and A. D. Singh, 1937, Recovery of Sulfur Dioxide from Waste Gases, Industrial and Engineering Chemistry, Vol. 29, No. 8, pp. 286-297

5. H. F. Johnstone, 1951, Gas Cleaning and Purification, In: Encyclopedia of Chemical Technology, Vol. 7, R. E. Kirk and D. F. Othmer, eds. The Interscience Encyclopedia, Inc., New York, pp. 63-106

6. H. F. Johnstone and G. C. Williams, 1939, Absorption of Gases by Liquid Droplets, Industrial and Engineering Chemistry, Vol. 31, No. 8, pp. 993-1001

7. H. F. Johnstone and M. H. Roberts, 1949, Deposition of Aerosol Particles from Moving Gas Streams, Industrial and Engineering Chemistry, Vol. 41, No. 11, pp. 2417-2423

8. H. F. Johnstone and A. D. Singh, 1940, Recovery of Sulfur Dioxide from Waste Gases, Industrial and Engineering Chemistry, Vol. 32, No. 8, pp. 1037-1049

9. R. L. Copson and J. W. Payne, 1933, Recovery of Sulfur Dioxide as Dilute Sulfuric Acid, Industrial and Engineering Chemistry, Vol. 25, No. 8, pp. 909-916

10. M. Noguchi, 1974, Status Report on Chiyoda Thoroughbred 101 Process, In: Proceedings: Symposium on Flue Gas Desulfurization - Atlanta, November 1974, Vol. II, EPA-650/2-74-126-b, U.S. Environmental Protection Agency, Washington, D.C., pp. 837-850

11. G. Nonhebel, 1936, A Commercial Plant for Removal of Smoke and Oxides of Sulphur from Flue Gases, Transactions of the Faraday Society, Vol. XXXII, pp. 1291-1297

12. A. V. Slack, 1971, Sulfur Dioxide Removal from Waste Gases, Noyes Data Corp., Park Ridge, New Jersey. See also Ref. 28

13. G. Nonhebel, 1981, Industrial Air Pollution: British Progress - A Review, Atmospheric Environment, Vol. 15, pp. 213-219

14. F. M. Kennedy and S. V. Tomlinson, 1978 , Flue Gas Desulfurization in the United States - 1977, App. F, 1977 Technology Status Report, Argonne National Laboratory, Argonne, Ill. Also: TVA Bulletin Y-125, U.S. Tennessee Valley Authority, Muscle Shoals, Alabama

15. R. H. Borgwardt, 1974, EPA/RTP Pilot Studies Related to Unsaturated Operation of Lime and Limestone Scrubbers, In: Proceedings: Symposium on Flue Gas Desulfurization - Atlanta, November 1974, Vol. I, EPA-650/2-74-126a, U.S. Environmental Protection Agency, Washington, D.C., pp. 225-240

16. Tennessee Valley Authority, 1973, Full-Scale Desulfurization of Stack Gas by Dry Limestone Injection, 3 volumes, EPA-650/2-73-019-a,b,c, U.S. Environmental Protection Agency, Washington, D.C.

17. See: 1972, Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, 2 volumes, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C. 1973, Proceedings: Flue Gas Desulfurization Symposium - 1973, EPA-650/2-73-038, U.S. Environmental Protection Agency, Washington, D.C.

18. This trend is clearly evident in a year-by-year comparison of process discussions in the proceedings of the early EPA FGD symposiums. See Ref. 17

19. P. G. Maurin and J. Jonikin, 1970, Removing Sulfur Oxides from Stacks, Chemical Engineering, Vol. 77, No. 9, pp. 173-180.

20. R. P. Hangebrauck and P. W. Spaite, 1968, Controlling the Oxides of Sulfur, Journal of the Air Pollution Control Association, Vol. 18, No. 1, pp. 5-8

21. P. S. Lowell, D. M. Ottmers, Jr., K. Schwitzgebel, T. I. Strange, and D. W. DeBerry, (Radian Corp.), 1970, A Theoretical Description of the Limestone Injection - Wet Scrubbing Process, Final report to the National Air Pollution Control Administration, NTIS PB 193029

22. M. Epstein, 1975, EPA Alkali Scrubbing Test Facility: Summary of Testing Through October 1974, EPA-650/2-75-047, U.S. Environmental Protection Agency, Washington, D.C.

23. M. Epstein, 1975, EPA Alkali Scrubbing Test Facility: Advanced Program, First Progress Report, EPA-600/2-75-050, U.S. Environmental Protection Agency, Washington, D.C.

24. H. N. Head, 1976, EPA Alkali Scrubbing Test Facility: Advanced Program, Second Progress Report, EPA-600/7-76-008, U.S. Environmental Protection Agency, Washington, D.C.

25. H. N. Head, 1977, EPA Alkali Scrubbing Test Facility: Advanced Program, Third Progress Report, EPA-600/7-77-105, U.S. Environmental Protection Agency, Washington, D.C.

26. H. N. Head and S.-C. Wang, 1979, EPA Alkali Scrubbing Test Facility: Advanced Program, Fourth Progress Report, 2 volumes, EPA-600/7-79-244a and -244b, U.S. Environmental Protection Agency, Washington, D.C.

27. D. A. Burbank and S.-C. Wang, 1980, EPA Alkali Scrubbing Test Facility: Advanced Program - Final Report (October 1974-June 1978), EPA-600/7-80-115, U.S. Environmental Protection Agency, Washington, D.C.

28. G. T. Rochelle, 1977, Process Synthesis and Innovation in Flue Gas Desulfurization, ph.D. Dissertation, University of California at Berkeley

29. G. T. Rochelle and C. J. King, 1977, The Effect of Additives on Mass Transfer in $CaCO_3$ or CaO Slurry Scrubbing of SO_2 from Waste Gases, Industrial and Engineering Chemistry, Fundamentals, Vol. 16, No. 1, pp. 67-75.

30. W. E. Corbett, O. W. Hargrove, and R. S. Merrill, (Radian Corp.), 1977, A Summary of the Effects of Important Chemical Variables Upon the Performance of Lime/Limestone Wet Scrubbing Systems, EPRI FP-639, Electric Power Research Institute, Palo Alto, Calif.

31. 1974, Langes Handbook of Chemistry, 11th Ed., J. A. Dean, ed., McGraw-Hill, New York.

32. M. Falk and P. A. Giguère, 1958, On the Nature of Sulfurous Acid, Canadian Journal of Chemistry, Vol. 36, pp. 1121-1125

33. M. Schmidt and W. Siebert, 1973, Sulphur, In: Comprehensive Inorganic Chemistry, Vol. 2, Pergamon Press, Oxford, pp. 795-933

34. A. R. Davis and R. M. Chatterjee, 1975, A Vibrational-Spectroscopic Study of the SO_2-H_2O System, Journal of Solution Chemistry, Vol. 4, No. 5, pp. 399-412

35. D. Benson, 1968, Mechanisms of Inorganic Reactions in Solution, McGraw-Hill, London

36. N. Kaplan, 1976, Introduction to Double Alkali Flue Gas Desulfurization, In: Proceedings: Symposium on Flue Gas Desulfurization - New Orleans, March 1976, Vol. I, EPA-600/2-76-136a, U.S. Environmental Protection Agency, Washington, D.C., pp. 387-421

37. F. A. Cotton and G. Wilkinson, 1966, Advanced Inorganic Chemistry, Interscience Publishers, New York

38. R. H. Borgwardt, 1976, IERL-RTP Scrubber Studies Related to Forced Oxidation, In: Proceedings: Symposium on Flue Gas Desulfurization, New Orleans, March 1976, Vol. I, EPA-600/2-76-136a, U.S. Environmental Protection Agency, Washington, D.C., pp. 117-143

39. Y. K. Kim, M. E. Deming, and J. D. Hatfield, 1975, Dissolution of Limestone in Simulated Slurries for Removal of Sulfur Dioxide from Stack Gases, Environmental Science & Technology, Vol. 9, No. 10, pp. 949-952

40. D. C. Drehmel, 1971, Limestone Types for Flue Gas Scrubbing, In: Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, Vol. I, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C., pp. 167-194

41. C.-y. Wen, W. J. McMichael, and R. D. Nelsen, Jr., 1975, Scale Control in Limestone Wet Scrubbing Systems, EPA-650/2-75-031, U.S. Environmental Protection Agency, Washington, D.C.

42. G. T. Paul, 1981, Corrosion in Flue Gas Scrubbers and Some Methods of Mitigation, Materials Performance, Vol. 20, No, 2, pp. 44-48

43. D. A. Hegg and P. V. Hobbs, 1978, Oxidation of Sulfur Dioxide in Aqueous Systems with Particular Reference to the Atmosphere, Atmospheric Environment, Vol. 12, pp. 241-253

44. T.-I. Chen and C. H. Barron, 1972, Some Aspects of the Homogeneous Kinetics of Sulfite Oxidation, Industrial and Engineering Chemistry, Fundamentals, Vol. II, No. 4, pp. 466-470

45. T. V. Larson, N. R. Horike, and H. Harrison, 1978, Oxidation of Sulfur Dioxide by Oxygen and Ozone in Aqueous Solution: A Kinetic Study with Significance to Atmospheric Rate Processes, Atmospheric Environment, Vol. 12, pp. 1597-1611

46. E. Hayon, A. Treinin, and J. Wilf, 1972, Electronic Spectra, Photochemistry, and Auto-oxidation Mechanism of the Sulfite-Bisulfite-Pyrosulfite Systems. The SO_2^-, SO_3^-, SO_4^-, and SO_5^- Radicals, Journal of the American Chemical Society, Vol. 94, No. 1, pp. 47-57

47. S. Beilke and G. Gravenhorst, 1978, Heterogeneous SO_2- Oxidation in the Droplet Phase, Atmospheric Environment, Vol. 12, pp. 231-239

48. J. H. Overton, Jr., V. P. Aneja, and J. L. Durham, 1979, Production of Sulfate in Rain and Raindrops in Polluted Atmospheres, Atmospheric Environment, Vol. 13, pp. 355-367

49. R. F. Adamowicz, 1979, A Model for the Reversible Washout of Sulfur Dioxide, Ammonia and Carbon Dioxide from a Polluted Atmosphere and the Production of Sulfates in Raindrops, Atmospheric Environment, Vol. 13, pp. 105-121

50. E. C. Fuller and R. H. Crist, 1941, The Rate of Oxidation of Sulfite Ions by Oxygen, Journal of the American Chemical Society, Vol. 63, pp. 1644-1650

51. K. J. A. de Wall and J. C. Okeson, 1966, The Oxidation of Aqueous Sodium Sulphite Solutions, Chemical Engineering Science, Vol. 21, pp. 559-572

52. V. Linek and J. Mayrhoferová, 1970, The Kinetics of Oxidation of Aqueous Sodium Sulphite Solution, Chemical Engineering Science, Vol. 25, pp. 787-800

53. J. L. Hudson, J. Erwin, and N. M. Catipovic, 1979, Kinetics of Sulfur Dioxide Oxidation in Aqueous Solution, EPA-600/7-79-030, U.S. Environmental Protection Agency, Washington, D.C.

54. H. F. Johnstone, 1931, Metallic Ions as Catalysts for the Removal of Sulfur Dioxide from Boiler Furnace Gases, Industrial and Engineering Chemistry, Vol. 23, No. 5, pp. 559-561

55. H. N. Alyea and H. L. Backstrom, 1929, The Inhibitive Action of Alcohols on the Oxidation of Sodium Sulfite, Journal of the American Chemical Society, Vol. 51, pp. 90-109

56. J. L. Hudson, 1980, Sulfur Dioxide Oxidation in Scrubber Systems, EPA-600/7-80-083, U.S. Environmental Protection Agency, Washington, D.C.

57. T. Nash, 1979, The Effect of Nitrogen Dioxide and of Some Transition Metals on the Oxidation of Dilute Bisulphite Solutions, Atmospheric Environment, Vol. 13, pp. 1149-1154. Discussion and reply, Vol. 14, p. 271

58. H. S. Rosenberg and H. M. Grotta, 1980, NO_x Influence on Sulfite Oxidation and Scaling in Lime/Limestone Flue Gas Desulfurization (FGD) Systems, Environmental Science & Technology, Vol. 14, No. 4, pp. 470-472

59. M. G. Manvelyan, G. O. Grigoryan, S. A. Gazaryan, G. S. Papyan, S. S. Karakhanyan, and G. A. Arutyunyan, 1961, Effect of Inhibitors on Oxidation of Magnesium Sulfite by Atmospheric Oxygen in Presence of Traces of Nitrogen Oxides, Journal of Applied Chemistry of the USSR, Vol. 34, pp. 896-898

60. G. Tarbutton, J. C. Driskell, T. M. Jones, F. J. Gray, and C. M. Smith, 1957, Recovery of Sulfur Dioxide from Flue Gases, Industrial and Engineering Chemistry, Vol. 49, No. 3, pp. 392-395

61. J. Ando, 1973, Status of Japanese Flue Gas Desulfurization Technology, In: Proceedings: Flue Gas Desulfurization Symposium - 1973, EPA-650/2-73-038, U.S. Environmental Protection Agency, Washington, D.C., pp. 69-101

62. Y. Yamamichi and J. Nagao, 1976, The Dowa's Basic Aluminum Sulfate-Gypsum Flue Gas Desulfurization Process, In: Proceedings: Symposium on Flue Gas Desulfurization - New Orleans, March 1976, Vol. II, EPA 600/2-76-136b, U.S. Environmental Protection Agency, Washington, D.C., pp. 833-842

63. T. Uno, M. Atsukawa, and K. Muramatsu, 1972, The Pilot Scale R&D and Prototype Plant of MHI Lime-Gypsum Process, In: Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, Vol. II, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C., pp. 833-849

64. R. W. Goodwin, 1978, Oxidation of Flue Gas Desulfurization Waste and the Effect on Treatment Modes, Journal of the Air Pollution Control Association, Vol. 28, No. 1, pp. 35-39

65. D. D. Clasen and H. Idemura, 1978, Limestone/Gypsum Jet Bubbling Scrubbing System, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, Fla., November 1977, Vol. II, F. A. Ayer, ed., EPA-600/7-78-058b, U.S. Environmental Protection Agency, Washington, D.C., pp. 837-856

66. G. P. Behrens and O. W. Hargrove, 1980, Evaluation of Chiyoda Thoroughbred 121 FGD Process and Gypsum Stacking, Vols. 1 and 2, EPRI CS-1579, Electric Power Research Institute, Palo Alto, California

67. R. J. Kruger, 1978, Experience with Limestone Scrubbing, Sherburne County Generating Plant, Northern States Power Company, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, Fla., November 1977, Vol. I, F. A. Ayer, ed., EPA-600/7-78-058a, U.S. Environmental Protection Agency, Washington, D.C., pp. 292-319

68. R. H. Borgwardt, 1977, Sludge Oxidation in Limestone FGD Scrubbers, EPA-600/7-77-061, U.S. Environmental Protection Agency, Washington, D.C.

69. R. H. Borgwardt, 1978, Effect of Forced Oxidation on Limestone/SO_x Scrubber Performance, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, Fla., November 1977, Vol. I, EPA-600/7-78-058a, U.S. Environmental Protection Agency, Washington, D.C., pp. 205-228.

70. I. J. Urza and M. L. Jackson, 1975, Pressure Aeration in a 55-Ft Bubble Column, Industrial and Engineering Chemistry, Process Design and Development, Vol. 14, No. 2, pp. 106-113

71. I. S. Shah, 1972, SO_2 Removal Using Calcium Based Alkalies, Pilot Plant Experience, In: Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, Vol. I, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C., pp. 345-372

72. J. M. Potts, A. V. Slack, and J. D. Hatfield, 1972, Removal of Sulfur Dioxide from Stack Gases by Scrubbing with Limestone Slurry: Small-Scale Studies at TVA, In: Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, New Orleans, Louisiana, November 8-12, 1971, Vol. I, U.S. Environmental Protection Agency, Washington, D.C., pp. 195-231

73. D. H. Stowe, D. S. Henzel, D. C. Hoffman, 1980, The FGD Dilemma; Lime, Limestone, or Thiosorbic Lime, In: Proceedings: Symposium on Flue Gas Desulfurization - Las Vegas, Nevada, March 1979, Vol. II, F. A. Ayer, ed., EPA-600/7-79-167b, U.S. Environmental Protection Agency, Washington, D.C., pp. 959-988

74. R. L. Nelson and R. D. O'Hara, 1977, Operating Experiences at the Phillips and Elrama Flue Gas Desulfurization Facilities, In: Proceedings: The Second Pacific Chemical Engineering Congress (Pachec '77), Vol. 1, American Institute of Chemical Engineers, New York, pp. 308-315

75. W. A. Cronkright and W. J. Leddy, 1976, Improving Mass Transfer Characteristics of Limestone Slurries by Use of Magnesium Sulfate, Environmental Science & Technology, Vol. 10, No. 6, pp. 569-572

76. J. G. Selmeczi, 1974, Separation of Sulfur Dioxide from Gas Stream, German Patent 2 421 423.

77. W. A. Cronkright, Jr., and B. G. Mandelik, 1975, Removing Sulfur-Containing Gases from Waste Gas, U.S. Patent 3 883 639.

78. J. D. Hatfield and J. M. Potts, 1972, Removal of Sulfur Dioxide from Stack Gases by Scrubbing with Limestone Slurry: Use of Organic Acids, In: Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, Vol. I, APTD-1161, U. S. Environmental Protection Agency, Washington, D.C., pp. 263-283

79. G. T. Rochelle and C. J. King, 1977, Process Alternates for Stack Gas Desulfurization by Throwaway Scrubbing, In: Proceedings: The Second Pacific Chemical Engineering Congress (Pachec '77), Vol. I, American Institute of Chemical Engineers, New York, pp. 261-267

80. H. N. Head, S.-C. Wang, D. T. Rabb, R. H. Borgwardt, J. E. Williams, and M. A. Maxwell, 1979, Recent Results from EPA's Lime/Limestone Scrubbing Programs - Adipic Acid as a Scrubber Additive, In: Proceedings: Symposium on Flue Gas Desulfurization - Las Vegas, Nevada, March 1979, Vol. I, F. A. Ayer, ed., EPA-600/7-79-167a, U.S. Environmental Protection Agency, Washington, D.C., pp. 342-385

81. D. A. Burbank, S.-C. Wang, R. R. McKinsey, and J. E. Williams, 1980, Test Results on Adipic Acid-Enhanced Limestone Scrubbing at the EPA Shawnee Test Facility - Third Report, In: Proceedings: Symposium on Flue Gas Desulfurization - Houston, October 1980, Vol. I, EPA 600/9-81-019a, U.S. Environmental Protection Agency, Washington, D.C., pp. 233-280

82. S.-C. Wang and D. A. Burbank, (Bechtel National, Inc.), 1981, Adipic Acid-Enhanced Lime/Limestone Test Results at the EPA Alkali Scrubbing Test Facility, preprint, paper presented at the EPA industry briefing on adipic acid enhanced limestone flue gas desulfurization process, Springfield, Missouri, July 15, 1981

83. O. W. Hargrove, J. D. Colley, G. E. Brown, 1981, Status of Adipic Acid Enhanced FGD System Demonstration at Springfield City Utilities Southwest Power Plant, preprint, paper presented at the EPA industry briefing on adipic acid enhanced limestone flue gas desulfurization process, Springfield, Missouri, July 15, 1981

84. R. W. Gerstle and D. S. Henzel, 1981, Evaluation of the Adipic Acid Enhanced Limestone Flue Gas Desulfurization Process on an Industrial Boiler, preprint, paper presented at the EPA industrial briefing on adipic acid enhanced limestone flue gas desulfurization process, Springfield, Missouri, July 15, 1981

85. J.C.S. Chang, 1981, Effect of Scrubbing Operating Conditions on Adipic Acid Degradation, EPA-600/7-81-017 (NTIS PB 81-163305), U.S. Environmental Protection Agency, Washington, D.C.

86. F. B. Meserole, D. L. Lewis, and F. T. Kurzawa, 1980, Further Study of Adipic Acid Degradation in FGD Scrubbers, EPA 600/PS-7-80-152, Project Summary, U.S. Environmental Protection Agency, Washington, D.C.

87. Extensive discussion of scale problems is found in papers of the early EPA FGD symposium proceedings. See: Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, 2 volumes, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C., in which a session was devoted to scale problems; and Proceedings: Flue Gas Desulfurization Symposium - 1973, EPA-650/2-73-038, U.S. Environmental Protection Agency, Washington, D.C.

88. S. Uchida, C.-y. Wen, and W. J. McMichael, 1976, Role of Holding Tank in Lime and Limestone Slurry Sulfur Dioxide Scrubbing, Industrial and Engineering Chemistry, Process Design and Development, Vol. 15, No. 1, pp. 88-95

89. J. W. Jones and R. D. Stern, 1973, Waste Products from Throwaway Flue Gas Cleaning Processes - Ecologically Sound Treatment and Disposal, In: Proceedings: Flue Gas Desulfurizatioin Symposium - 1973, EPA 650/2-73-038, U.S. Environmental Protection Agency, Washington, D.C., pp. 187-234

90. M. P. Smith, M. T. Melia, B. A. Laseke, Jr., and N. Kaplan, 1980, Recent Trends in Flue Gas Desulfurization, In: Proceedings: Symposium on Flue Gas Desulfurization - Houston, October, Vol. I, pp. 115-172

91. Versar, Inc., 1979, Selection of Representative Coal Ash and Coal Ash/FGD Waste Disposal Sites for Future Testing, draft report, U.S. Environmental Protection Agency, Washington, D.C.

92. R. G. Knight, E. H. Rothfuss, and K. D. Yard, (Michael Baker Jr., Inc.), 1980, FGD Sludge Disposal Manual, 2nd Ed., EPRI CS-1515, Electric Power Research Institute, Palo Alto, Calif.

93. J. Rossoff, R. C. Rossi, R. B. Fling, W. M. Graven, and P. P. Leo, (The Aerospace Corp.), 1979, Disposal of Byproducts from Nonregenerable Flue Gas Desulfurization Systems: Final Report, EPA-600/7-79-046, U.S. Environmental Protection Agency, Washington, D.C.

94. D. E. Weaver, C. J. Schmidt, and J. P. Woodyard, (SCS Engineers), 1977, Data Base for Standards/Regulations Development for Land Disposal of Flue Gas Cleaning Sludges, EPA-600/7-77-118, U.S. Environmental Protection Agency, Washington, D.C.

95. P. P. Leo, R. B. Fling, and J. Rossoff, 1978, Flue Gas Desulfurization Waste Disposal Field Study at the Shawnee Power Station, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977, Vol. II, F. A. Ayer, ed., EPA-600/7-78-058b, U.S. Environmental Protection Agency, Washington, D.C., pp. 496-536

96. J. W. Barrier, H. L. Faucett, and L. J. Henson, 1978, Economics of Disposal of Lime/Limestone Scrubbing Wastes: Untreated and Chemically Treated Wastes, EPA-600/7-78-023a, U.S. Environmental Protection Agency, Washington, D.C.

97. ______, 1979, Economics of Disposal of Lime/Limestone Scrubbing Wastes: Sludge/Flyash Blending and Gypsum Processes, EPA-600/7-79-069, U.S. Environmental Protection Agency, Washington, D.C.

98. J. D. Veitch, A. E. Steele, and T. W. Tarkington, 1980, Economics of Disposal of Lime/Limestone Scrubbing Wastes: Surface Mine Disposal and Dravo Landfill Processes, EPA-600/7-80-022, U.S. Environmental Protection Agency, Washington, D.C.

99. C. J. Santhanam, R. R. Lunt, C. B. Cooper, D. E. Klimschmidt, I. Bodek, W. A. Tucker, and C. R. Ullrich, 1980, Waste and Water Management for Conventional Coal Combustion, Assessment Report - 1979, Volume V. Disposal of FGC Wastes, EPA-600/7-80-012e, U.S. Environmental Protection Agency, Washington, D.C.

100. W. A. Duvel, Jr., W. R. Gallagher, R. G. Knight, C. R. Kolarz, R. J. McLaren, 1978, State-of-the-Art of FGD Sludge Fixation, EPRI FP-671, Electric Power Research Institute, Palo Alto, Calif.

101. W. M. Coltharp and others, (Radian Corp.), 1979, Review and Assessment of the Existing Data Base Regarding Flue Gas Cleaning Wastes, EPRI FP-671, Vol. 1, Electric Power Research Institute, Palo Alto, Calif.

102. W. M. Coltharp and others, (Radian Corp.), 1979, Chemical/Physical Stability of Flue Gas Cleaning Wastes, EPRI FP-671, Vol. 2, Electric Power Research Institute, Palo Alto, Calif.

103. B. F. Jones, C. M. Thompson, A. G. Lamkin, and K. R. Williams, 1980, Studies of Long-Term Chemical and Physical Properties of Mixtures of Flue Gas Cleaning Wastes, EPRI CS-1533, Electric Power Research Institute, Palo Alto, Calif.

104. J. H. Wilhelm, R. W. Kobler, Y. Naide, and G. Redfield, (Envirotech Corp.), 1979, Sludge Dewatering for FGD Products, EPRI FP-937, Electric Power Research Institute, Palo Alto, Calif.

105. 1974, Process Design Manual for Sludge Treatment and Disposal, EPA-625/1-74-006, U.S. Environmental Protection Agency, Washington, D.C.

106. D. J. Hagerty, C. R. Ullrich, and B. K. Thacker, 1977, Engineering Properties of FGD Sludges, In: Proceedings of the Conference on Geotechnical Practices for Disposal of Solid Waste Materials, American Society of Civil Engineers, New York, pp. 23-40

107. R. J. Bacskal and L. C. Cleveland, 1979, Design and Commercial Operation of Lime/Limestone FGD Sludge Stabilization Systems, In: Proceedings: Symposium on Flue Gas Desulfurization - Las Vegas, Nevada, March 1979, Vol. II, F. A. Ayer, ed., EPA-600/7-79-167b, U.S. Environmental Protection Agency, Washington, D.C., pp. 792-808

108. C. J. Santhanam, R. R. Lunt, and C. B. Cooper, 1979, Current Alternatives for Flue Gas Desulfurization (FGD) Waste Disposal - An Assessment, In: Proceedings: Symposium on Flue Gas Desulfurization - Las Vegas, Nevada, March 1979, Vol. I, F. A. Ayer, ed., EPA-600/7-79-167a, U.S. Environmental Protection Agency, Washington, D.C., pp. 561-594

109. J. Rossoff, R. C. Rossi, R. B. Fling, W. M. Graven, and P. P. Leo, (The Aerospace Corp.), 1977, Disposal of By-Products From Nonregenerable Flue Gas Desulfurization Systems: Second Progress Report, EPA-600/7-77-052, U.S. Environmental Protection Agency, Washington, D.C.

110. J. E. Garlanger and T. S. Ingra, (Ardaman Associates, Inc.), 1980, Evaluation of Chiyoda Thoroughbred 121 FGD Process and Gypsum Stacking, Volume 3: Testing the Feasibility of Stacking FGD Gypsum, EPRI CS-1579, Electric Power Research Institute, Palo Alto, Calif. See also, Volume 3 Addendum, 1981

111. M. R. Gogineni, J. R. Martin, and P. G. Maurin, 1973, Status of C-E's Air Quality Control Systems, In: Proceedings: Flue Gas Desulfurization Symposium - 1973, EPA-650/2-73-038, U.S. Environmental Protection Agency, Washington, D.C., pp. 539-551

112. W. P. Jones, 1949, Development of the Venturi Scrubber, Industrial and Engineering Chemistry, Vol. 41, No. 11, pp. 2424-2427

113. D. M. Muir and Y. Miheisi, 1979, Comparison of Performance of a Single and Two-Stage Variable-Throat Venturi Scrubber, Atmospheric Environment, Vol. 13, pp. 1187-1196

114. K. C. Goel and K. G. T. Hollands, 1977, A General Method for Predicting Particulate Collection Efficiency of Venturi Scrubbers, Industrial and Engineering Chemistry, Fundamentals, Vol. 16, No. 2, pp. 186-193

115. S. Calvert, 1970, Venturi and Other Atomizing Scrubbers, Efficiency and Pressure Drop, American Institute of Chemical Engineers Journal, Vol. 16, No. 3, pp. 392-396

116. R. H. Boll, 1973, Particle Collection and Pressure Drop in Venturi Scrubbers, Industrial and Engineering Chemistry, Fundamentals, Vol. 12, No. 1, pp. 40-50

117. K. G. T. Hollands and K. C. Goel, 1975, A General Method for Predicting Pressure Loss in Venturi Scrubbers, Industrial and Engineering Chemistry, Fundamentals, Vol. 14, No. 1, pp. 16-22

118. S. W. Behie and J. M. Beeckmans, 1973, On the Efficiency of a Venturi Scrubber, The Canadian Journal of Chemical Engineering, Vol. 51, pp. 430-433

119. C. P. Kerr, 1974, Sulfur Dioxide Removal in Venturi Scrubbers, Industrial and Engineering Chemistry, Process Design and Development, Vol. 13, No. 3, pp. 222-225

120. S. Uchida and C.-y. Wen, 1973, Gas Absorption by Alkaline Solutions in a Venturi Scrubber, Industrial and Engineering Chemistry, Process Design and Development, Vol. 12, No. 4, pp. 437-443

121. 1973, Chemical Engineers' Handbook, 5th Ed., McGraw-Hill, New York

122. A. Saleem, 1980, Spray Tower: The Workhorse of flue-gas desulfurization, Power, Vol. 124, No. 10, pp. 73-77

123. 1963, New Approach Used in Liquid-Gas Contactor, Chemical and Engineering News, Vol. 41, No. 34, pp. 44-45

124. P. Sorenson, N. E. Takvoryan, and R. J. Jaworowski, 1976, TCA Sphere Development and Evaluation, In: Proceedings: Symposium on Flue Gas Desulfurization - New Orleans, March 1976, Vol. II, EPA-600/2-76-136b, U.S. Environmental Protection Agency, Washington, D.C., pp. 999-1015

125. S. Uchida, C.-s. Chang, and C.-y. Wen, 1977, Mechanics of a Turbulent Contact Absorber, The Canadian Journal of Chemical Engineering, Vol. 55, pp. 392-396

126. J. Tichy, A. Wong, and W. J. M. Douglas, 1972, Pressure Drop in a Mobile-Bed Contactor, The Canadian Journal of Chemical Engineering, Vol. 50, pp. 215-220

127. J. Tichy and W. J. M. Douglas, 1972, Bed Expansion in a Mobile-Bed Contactor, The Canadian Journal of Chemical Engineering, Vol. 50, pp. 702-707

128. B. H. Chen and W. J. M. Douglas, 1968, Liquid Hold-up and Minimum Fluidization Velocity in a Turbulent Contactor, The Canadian Journal of Chemical Engineering, Vol. 46, pp. 245-249

129. C.-y. Wen and L. S. Fan, 1975, Absorption of Sulfur Dioxide in Spray Column and Turbulent Contacting Absorbers, EPA-600-75-023 (NTIS PB 247334), U.S. Environmental Protection Agency, Washington, D.C.

130. E. F. Aul, Jr., R. D. Delleney, G. D. Brown, G. C. Page, and D. O. Stuebner, (Radian Corp.), 1976, Evaluation of Regenerable Flue Gas Desulfurization Processes, EPRI FP-272, Vol. II, Electric Power Research Institute, Palo Alto, Calif., pp. 335-375

131. Although individually specific, the many discussions in the proceedings of the early EPA FGD symposiums provide, in aggregate, a comprehensive representation. See: 1972, Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, 2 volumes, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C. 1973, Proceedings: Flue Gas Desulfurization Symposium - 1973, EPA-650/2-73-038, U.S. Environmental Protection Agency, Washington, D.C. 1974, Proceedings: Symposium on Flue Gas Desulfurization - Atlanta, November 1974, EPA-650/2-74-126-a and -126-b, U.S. Environmental Protection Agency, Washington, D.C.

132. J. Javetski, 1978, *Solving Corrosion Problems in Air-Pollution Control Equipment - Part 1*, Power, Vol. 122, No. 5, pp. 72-77; *... - Part 2*, Vol. 122, No. 6, pp. 80-87

133. E. C. Hoxie and G. W. Tuffnell, 1976, *A Summary of Corrosion Tests in Flue Gas Desulfurization Processes*, Journal of the Air Pollution Control Association, Vol. 26, No. 4, pp. 307-312

134. F. G. Hodge, 1978, *High Performance Alloys ... Make Wet Scrubbers Work*, Chemical Engineering Progress, Vol. 74, No. 10, pp. 84-88

135. The periodical *Materials Performance*, published by the National Association of Corrosion Engineers, contains many substantial discussions of FGD corrosion topics

136. Most descriptions of full-scale systems in the proceedings of the EPA FGD symposiums discuss corrosion-related problems and their solutions. See Ref. 17 and 1976, *Proceedings: Symposium on Flue Gas Desulfurization, New Orleans, March 1976*, EPA-600/2-76-136a and -136b, U.S. Environmental Protection Agency, Washington, D.C. 1978, *Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977*, F. A. Ayer, ed., EPA-600/7-78-058a and -058b, U.S. Environmental Protection Agency, Washington, D.C.

137. G. L. Crowe and H. R. Horseman, 1976, *Third TVA Interim Report of Corrosion Studies: EPA Alkali Scrubbing Test Facility*, Ref. 24, Appendix L. _____, 1977, *Fourth TVA Interim Report of Corrosion Studies: EPA Alkali Scrubbing Test Facility*, Ref. 25, Appendix L. _____, 1979, *Fifth Interim Report of Corrosion Studies: EPA Alkali Scrubbing Test Facility*, Ref. 26, Appendix K

138. P. S. Maiya, 1980, *A State-of-the-Art Review of Materials-Related Problems in Flue Gas Desulfurization Systems*, ANL-80-59, Argonne National Laboratory, Argonne, Ill.

139. H. S. Rosenberg and others (Battelle, Columbus Laboratories), 1981, *Construction Materials for Wet Scrubbers*, 2 volumes, EPRI CS-1736, Electric Power Research Institute, Palo Alto, Calif.

140. G. Paul, 1978, *Corrosion Resistant Materials for SO_2 Scrubbers*, Power Engineering, Vol. 82, No. 5, pp. 54-57

141. N. T. Mistry, 1976, *Material Selection for Gas Scrubbers*, Materials Performance, Vol. 15, No. 4, pp. 27-33.

142. J. R. Crum, E. L. Hibner, and R. W. Ross, Jr., 1981, *Corrosion Resistance of High-Nickel Alloys in Simulated SO_2-Scrubber Environments*, Materials Performance, Vol. 20, No. 2, pp. 9-13

143. C. G. Munger, 1980, Practical Aspects of Coating Repair, Materials Performance, Vol. 19, No. 2, pp. 46-52

144. D. M. Berger, R. J. Trewella, and C. J. Wummer, 1980, Atlas Test Cell Evaluation of Coatings for SO_2 Scrubber Service, Materials Performance, Vol. 19, No. 4, pp. 25-28

145. W. L. Anders and R. L. Torstrick, 1981, Computerized Shawnee Lime/Limestone Scrubbing Model Users Manual, EPA-600/8-81-008, U.S. Environmental Protection Agency, Washington, D.C.

146. M. Epstein, C. C. Leivo, and C. H. Rowland, 1972, Mathematical Models for Pressure Drop, Particulate Removal and SO_2 Removal in Venturi, TCA, and Hydro-filter Scrubbers, In: Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, Vol. I, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C., pp. 45-113

147. M. Epstein, C. C. Leivo, C. H. Rowland, and S. C. Wang, 1973, Test Results from the EPA Lime/Limestone Scrubbing Test Facility, In: Proceedings: Flue Gas Desulfurization Symposium - 1973, EPA-650/2-73-038, U.S. Environmental Protection Agency, Washington, D.C., pp. 235-332

148. R. L. Torstrick, L. J. Henson, and S. V. Tomlinson, 1978, Economic Evaluation Techniques, Results, and Computer Modeling for Flue Gas Desulfurization, In: Proceedings: Symposium on Flue Gas Desulfurization - Hollywood, FL, November 1977, Vol. I, F. A. Ayer, ed., EPA-600/7-78-058a, U.S. Environmental Protection Agency, Washington, D.C., pp. 118-168

149. C. D. Stephenson and R. L. Torstrick, 1979, The Shawnee Lime-Limestone Computer Program, preprint, presented at the Fifth Industrial Briefing Conference, sponsored by the U.S. Environmental Protection Agency, at Raleigh, N. Car., December 5, 1979

150. D. M. Ottmers, Jr., 1972, A Model for the Limestone Injection - Wet Scrubbing Process for Sulfur Dioxide Removal from Power Plant Flue Gas, In: Proceedings: Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, Vol. I, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C., pp. 115-149

151. A. V. Slack and G. A. Hollinden, 1975, Sulfur Dioxide Removal From Waste Gas, 2nd Ed., Noyes Data Corporation, Park Ridge, New Jersey

152. F. A. Zenz, 1963, Absorption, In: Kirk-Othmer Encyclopedia of Chemical Technology, 2d Ed., Vol. 1, pp. 44-77

153. P. V. Danckwerts, 1970, Gas-Liquid Reactions, McGraw-Hill, New York

154. C. S. Chang and G. T. Rochelle, 1981, SO_2 Absorption Into Aqueous Solutions, American Institute of Chemical Engineers Journal, Vol. 27, No. 2, pp. 292-298

155. W. J. McMichael, L. S. Fan, and C.-y. Wen, 1976, Analysis of Sulfur Dioxide Wet Limestone Scrubbing Data from Pilot Plant Spray and TCA Scrubbers, Industrial and Engineering Chemistry, Process Design and Development, Vol. 15, No. 3, pp. 459-467

156. C.-y. Wen and C.-s. Chang, 1978, Absorption of SO_2 in Lime and Limestone Slurry: Pressure Drop Effect on Turbulent Contact Absorber Performance, Environmental Science & Technology, Vol. 12, No. 6, pp. 703-707

157. C.-y. Wen and S. Uchida, 1971, Simulation of SO_2 Absorption in a Venturi Scrubber, In: Proceedings: Second International Lime/Limestone Wet-Scrubbing Symposium, November 8-12, 1971, New Orleans, Louisiana, Vol. I, APTD-1161, U.S. Environmental Protection Agency, Washington, D.C., pp. 3-44

158. Federal Register, 1979, New Stationary Sources Performance Standards; Electric Utility Steam Generating Units, Vol. 44, No. 113, June 11, pp. 33580-33624

159. J. A. Cavallaro, M. J. Johnson, and A. W. Deubrouck, 1976, Sulfur Reduction Potential of the Coals of the United States, Bureau of Mines Report of Investigation RI 8118, U.S. Bureau of Mines, Washington. D.C.

160. J. W. Hamersima and M. L. Kraft, 1975, Applicability of the Meyers Process for Chemical Desulfurization of Coal: Survey of Thirty-Five Coals. EPA-650/2-74-025-A, U.S. Environmental Protection Agency, Washington, D.C.

161. National Electric Reliability Council, 1980, 1980 Summary of Projected Peak Demand, Generating Capability, and Fossil Fuel Requirements, National Electric Reliability Council, Princeton, New Jersey

162. National Coal Association, 1979, Steam-Electric Plant Factors, 1979, National Coal Association, Washington, D.C.

163. G. R. Fryling, 1966, Combustion Engineering, 2nd Ed., Combustion Engineering, Inc., New York

164. Babcock & Wilcox, 1975, Steam/Its Generation and Use, Babcock & Wilcox Co., New York

165. G. D. Friedlander, 1980, Sixteenth Steam Station Design Survey, Electrical World, Vol. 194, No. 8, November, pp. 67-82

166. Department of Energy, 1978, Steam-Electric Plant Construction Cost and Annual Production Expenses 1977, DOE/EIA-0033/3 (77), U.S. Department of Energy, Washington, D.C.; DOE, 1979, Steam-Electric Plant Air and Water Quality Control Data, for the Year Ended December 31, 1976, DOE/FERC 0036, U.S. Department of Energy, Washington, D.C. These are issued annually.

167. Federal Register, 1971, Standards of Performance for New Stationary Sources, Vol. 36, No. 247, December 23, pp. 24876-24895

168. Chemical Engineering, 1976, 1977, 1978, 1979, Economic Indicators, Vols. 83, 84, 85, and 86

169. V. W. Uhl, 1979, A Standard Procedure for Cost Analysis of Pollution Control Operations, Vols. I and II, EPA-600/8-79-018a and -018b, U.S. Environmental Protection Agency, Research Triangle Park, North Carolina

170. The Richardson Rapid System, Process Plant Estimation Standards, Vols. I, III, IV, 1978-1979 Ed. Richardson Engineering Services, Inc., Solano Beach, Calif.

171. EPRI, 1978, Technical Assessment Guide, EPRI PS-866-SR, Special Report, Electric Power Research Institute, Palo Alto, Calif.

172. E. L. Grant and W. G. Ireson, 1970, Principles of Engineering Economy, Ronald Press, New York

173. P. H. Jeynes, 1968, Profitability and Economic Choice, 1st Ed. The Iowa State University Press, Ames, Iowa

174. O. W. Hargrove, Jr., J. D. Colley, and J. David Mobley, 1981, Adipic-Acid-Enhanced Limestone Flue Gas Desulfurization System Commercial Demonstration, preprint, presented at the APCA/ASME Environmental Information Exchange, at Research Triangle Park, North Carolina, December 8-9, 1981.

175. Tenneesee Valley Authority, 1969, Sulfur Oxide Removal From Power Plant Stack Gas, PB 178-972, National Technical Information Service, Springfield, Virginia

176. G. G. McGlamery, R. L. Torstrick, W. J. Broadfoot, J. P. Simpson, L. J. Henson, S. V. Tomlinson, and J. F. Young, 1975, Detailed Cost Estimates for Advanced Effluent Desulfurization Processes, EPA-600/2-75-006, U.S. Environmental Protection Agency, Washington, D.C.

APPENDIX A

LIMESTONE PROCESS WITHOUT FORCED OXIDATION OR ADDITIVES

TABLE A-1. LIMESTONE PROCESS, CASE 1, WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(200-MW, 3.5% sulfur, 1979 NSPS, spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,283
Feed preparation	2,746
Gas handling	6,007
SO_2 absorption	10,794
Reheat	1,934
Solids separation	2,050
Total process capital	24,814
Services, utilities, and miscellaneous	1,489
Total direct investment excluding pond	26,303
Pond construction	7,726
Total direct investment	34,029
Indirect Investment	
Engineering design and supervision	2,104
Architect and engineering contractor	789
Construction expense	4,735
Contractor fees	1,578
Contingency	3,551
Waste disposal indirect investment	2,302
Total fixed investment	49,088
Other Capital Investment	
Allowance for startup and modifications	3,125
Interest during construction	7,658
Royalties	0
Land	1,017
Working capital	1,827
Total capital investment	62,715
Dollars per kW of generation capacity	313.57

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid 1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE A-2. LIMESTONE PROCESS, CASE 1,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, 3.5% sulfur, 1979 NSPS, spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	58.3 ktons	8.50/ton	496
Total raw materials cost			496
Conversion costs			
Operating labor and supervision			
FGD	20,280 man-hr	15.00/man-hr	304
Utilities			
Steam	221,600 klb	2.50/klb	554
Process water	110,800 kgal	0.14/kgal	16
Electricity	21,465,000 kWh	0.037/kWh	794
Maintenance			
Labor and material			2,599
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			4,336
Total direct costs			4,832
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,783
Total first-year operating and maintenance costs			6,615
Levelized capital charges (14.7% of total capital investment)			9,219
Total first-year annual revenue requirements			15,834
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			12,477
Levelized capital charges (14.7% of total capital investment)			9,219
Levelized annual revenue requirements			21,696

	M$	Mills/kWh
First-year annual revenue requirements	15.83	14.39
Levelized annual revenue requirements	21.70	19.72

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE A-3. LIMESTONE PROCESS, CASE 2, WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 0.7% sulfur, 1979 NSPS, spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,265
Feed preparation	2,364
Gas handling	11,134
SO_2 absorption	17,997
Reheat	0
Solids separation	1,686
Total process capital	34,446
Services, utilities, and miscellaneous	2,067
Total direct investment excluding pond	36,513
Pond construction	4,084
Total direct investment	40,597
Indirect Investment	
Engineering design and supervision	2,556
Architect and engineering contractor	730
Construction expense	5,842
Contractor fees	1,826
Contingency	4,747
Waste disposal indirect investment	1,127
Total fixed investment	57,425
Other Capital Investment	
Allowance for startup and modifications	4,177
Interest during construction	8,958
Royalties	0
Land	554
Working capital	2,121
Total capital investment	73,235
Dollars per kW of generation capacity	146.47

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE A-4. LIMESTONE PROCESS, CASE 2,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 0.7% sulfur, 1979 NSPS, spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	21.9 ktons	8.50/ton	186
Total raw materials cost			186
Conversion costs			
Operating labor and supervision			
FGD	26,430 man-hr	15.00/man-hr	396
Utilities			
Process water	151,000 kgal	0.14/kgal	21
Electricity	39,566,000 kWh	0.037/kWh	1,464
Maintenance			
Labor and material			3,044
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			4,994
Total direct costs			5,180
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,106
Total first-year operating and maintenance costs			7,286
Levelized capital charges (14.7% of total capital investment)			10,765
Total first-year annual revenue requirements			18,051
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			13,741
Levelized capital charges (14.7% of total capital investment)			10,765
Levelized annual revenue requirements			24,506

	M$	Mills/kWh
First-year annual revenue requirements	18.05	6.56
Levelized annual revenue requirements	24.51	8.91

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE A-5. LIMESTONE PROCESS, CASE 3,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 2.0% sulfur, 1979 NSPS, spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,310
Feed preparation	3,157
Gas handling	10,868
SO_2 absorption	17,710
Reheat	3,612
Solids separation	2,175
Total process capital	38,832
Services, utilities, and miscellaneous	2,330
Total direct investment excluding pond	41,162
Pond construction	9,280
Total direct investment	50,442
Indirect Investment	
Engineering design and supervision	2,881
Architect and engineering contractor	823
Construction expense	6,586
Contractor fees	2,058
Contingency	5,351
Waste disposal indirect investment	2,561
Total fixed investment	70,702
Other Capital Investment	
Allowance for startup and modifications	4,709
Interest during construction	11,030
Royalties	0
Land	1,217
Working capital	2,735
Total capital investment	90,393
Dollars per kW of generation capacity	180.79

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE A-6. LIMESTONE PROCESS, CASE 3, WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 2.0% sulfur, 1979 NSPS, spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	74.6 ktons	8.50/ton	634
Total raw materials cost			634
Conversion costs			
Operating labor and supervision			
FGD	30,680 man-hr	15.00/man-hr	460
Utilities			
Steam	551,100 klb	2.50/klb	1,378
Process water	216,400 kgal	0.14/kgal	30
Electricity	35,918,000 kWh	0.037/kWh	1,329
Maintenance			
Labor and material			3,571
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			6,872
Total direct costs			7,506
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,482
Total first-year operating and maintenance costs			9,988
Levelized capital charges (14.7% of total capital investment)			13,288
Total first-year annual revenue requirements			23,276
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			18,839
Levelized capital charges (14.7% of total capital investment)			13,288
Levelized annual revenue requirements			32,127

	M$	Mills/kWh
First-year annual revenue requirements	23.28	8.46
Levelized annual revenue requirements	32.13	11.68

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE A-7. LIMESTONE PROCESS, CASE 4,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,528
Feed preparation	4,715
Gas handling	11,320
SO_2 absorption	20,328
Reheat	3,635
Solids separation	2,646
Total process capital	45,172
Services, utilities, and miscellaneous	2,710
Total direct investment excluding pond	47,882
Pond construction	15,296
Total direct investment	63,178
Indirect Investment	
Engineering design and supervision	3,352
Architect and engineering contractor	958
Construction expense	7,661
Contractor fees	2,394
Contingency	6,225
Waste disposal indirect investment	4,222
Total fixed investment	87,990
Other Capital Investment	
Allowance for startup and modifications	5,478
Interest during construction	13,726
Royalties	0
Land	1,983
Working capital	3,375
Total capital investment	112,552
Dollars per kW of generation capacity	225.10

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE A-8. LIMESTONE PROCESS, CASE 4,
WITHOUT FORCED OXIDATION OR ADDITIVES
SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	142.9 ktons	8.50/ton	1,215
Total raw materials cost			1,215
Conversion costs			
Operating labor and supervision			
FGD	30,680 man-hr	15.00/man-hr	460
Utilities			
Steam	542,500 klb	2.50/klb	1,356
Process water	268,200 kgal	0.14/kgal	38
Electricity	50,623,000 kWh	0.037/kWh	1,873
Maintenance			
Labor and material			4,289
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			8,120
Total direct costs			9,335
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,912
Total first-year operating and maintenance costs			12,247
Levelized capital charges (14.7% of total capital investment)			16,545
Total first-year annual revenue requirements			28,792
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			23,097
Levelized capital charges (14.7% of total capital investment)			16,545
Levelized annual revenue requirements			39,642

	M$	Mills/kWh
First-year annual revenue requirements	28.79	10.47
Levelized annual revenue requirements	39.64	14.42

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE A-9. LIMESTONE PROCESS, CASE 5,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 5.0% sulfur, 1979 NSPS, spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,825
Feed preparation	5,326
Gas handling	11,362
SO_2 absorption	20,533
Reheat	3,606
Solids separation	2,949
Total process capital	46,601
Services, utilities, and miscellaneous	2,796
Total direct investment excluding pond	49,397
Pond construction	20,165
Total direct investment	69,562
Indirect Investment	
Engineering design and supervision	3,458
Architect and engineering contractor	988
Construction expense	7,904
Contractor fees	2,470
Contingency	6,422
Waste disposal indirect investment	5,566
Total fixed investment	96,370
Other Capital Investment	
Allowance for startup and modifications	5,650
Interest during construction	15,033
Royaltier	0
Land	2,603
Working capital	3,679
Total capital investment	123,335
Dollars per kW of generation capacity	246.67

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE A-10. LIMESTONE PROCESS, CASE 5,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 5.0% sulfur, 1979 NSPS, spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	204.3 ktons	8.50/ton	1,737
Total raw materials cost			1,737
Conversion costs			
Operating labor and supervision			
FGD	30,680 man-hr	15.00/man-hr	460
Utilities			
Steam	534,100 klb	2.50/klb	1,335
Process water	314,000 kgal	0.14/kgal	44
Electricity	54,388,000 kWh	0.037/kWh	2,012
Maintenance			
Labor and material			4,557
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			8,512
Total direct costs			10,249
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,072
Total first-year operating and maintenance costs			13,321
Levelized capital charges (14.7% of total capital investment)			18,130
Total first-year annual revenue requirements			31,451
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			25,123
Levelized capital charges (14.7% of total capital investment)			18,130
Levelized annual revenue requirements			43,253

	M$	Mills/kWh
First-year annual revenue requirements	31.45	11.44
Levelized annual revenue requirements	43.25	15.73

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE A-11. LIMESTONE PROCESS, CASE 6, WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, TCA, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,529
Feed preparation	4,761
Gas handling	12,101
SO_2 absorption	17,043
Reheat	3,476
Solids separation	2,625
Total process capital	42,535
Services, utilities, and miscellaneous	2,552
Total direct investment excluding pond	45,087
Pond construction	15,494
Total direct investment	60,581
Indirect Investment	
Engineering design and supervision	3,156
Architect and engineering contractor	902
Construction expense	7,214
Contractor fees	2,254
Contingency	5,861
Waste disposal indirect investment	4,276
Total fixed investment	84,244
Other Capital Investment	
Allowance for startup and modifications	5,158
Interest during construction	13,142
Royalties	0
Land	2,007
Working capital	3,234
Total capital investment	107,785
Dollars per kW of generation capacity	215.57

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE A-12. LIMESTONE PROCESS, CASE 6, WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, TCA, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	145.9 ktons	8.50/ton	1,240
Total raw materials cost			1,240
Conversion costs			
Operating labor and supervision			
FGD	30,680 man-hr	15.00/man-hr	460
Utilities			
Steam	507,100 klb	2.50/klb	1,268
Process water	270,000 kgal	0.14/kgal	38
Electricity	48,046,000 kWh	0.037/kWh	1,778
Maintenance			
Labor and material			4,071
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			7,719
Total direct costs			8,959
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,782
Total first-year operating and maintenance costs			11,741
Levelized capital charges (14.7% of total capital investment)			15,845
Total first-year annual revenue requirements			27,586
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			22,143
Levelized capital charges (14.7% of total capital investment)			15,845
Levelized annual revenue requirements			37,988

	M$	Mills/kWh
First-year annual revenue requirements	27.59	10.03
Levelized annual revenue requirements	37.99	13.81

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE A-13. LIMESTONE PROCESS, CASE 7,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,518
Feed preparation	4,617
Gas handling	13,406
SO_2 absorption	23,790
Reheat	3,537
Solids separation	2,600
Total process capital	50,468
Services, utilities, and miscellaneous	3,028
Total direct investment excluding pond	53,456
Pond construction	14,641
Total direct investment	68,137
Indirect Investment	
Engineering design and supervision	3,745
Architect and engineering contractor	1,070
Construction expense	8,559
Contractor fees	2,675
Contingency	6,954
Waste disposal indirect investment	4,041
Total fixed investment	95,181
Other Capital Investment	
Allowance for startup and modifications	6,120
Interest during construction	14,848
Royalties	0
Land	1,890
Working capital	3,670
Total capital investment	121,709
Dollars per kW of generation capacity	243.42

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE A-14. LIMESTONE PROCESS, CASE 7, WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	132.6 ktons	8.50/ton	1,127
Total raw materials cost			1,127
Conversion costs			
Operating labor and supervision			
FGD	32,320 man-hr	15.00/man-hr	485
Utilities			
Steam	493,100 klb	2.50/klb	1,233
Process water	262,300 kgal	0.14/kgal	36
Electricity	67,392,000 kWh	0.037/kWh	2,494
Maintenance			
Labor and material			4,719
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			9,071
Total direct costs			10,198
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,185
Total first-year operating and maintenance costs			13,383
Levelized capital charges (14.7% of total capital investment)			17,891
Total first-year annual revenue requirements			31,274
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			25,239
Levelized capital charges (14.7% of total capital investment)			17,891
Levelized annual revenue requirements			43,130

	M$	Mills/kWh
First-year annual revenue requirements	31.27	11.37
Levelized annual revenue requirements	43.13	15.68

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE A-15. LIMESTONE PROCESS, CASE 8, WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(1000-MW, 3.5% sulfur, 1979 NSPS, spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,915
Feed preparation	6,041
Gas handling	21,808
SO_2 absorption	39,922
Reheat	7,151
Solids separation	3,281
Total process capital	81,118
Services, utilities, and miscellaneous	4,867
Total direct investment excluding pond	85,985
Pond construction	25,433
Total direct investment	111,418
Indirect Investment	
Engineering design and supervision	5,159
Architect and engineering contractor	860
Construction expense	12,038
Contractor fees	3,439
Contingency	10,748
Waste disposal indirect investment	6,460
Total fixed investment	150,122
Other Capital Investment	
Allowance for startup and modifications	9,458
Interest during construction	23,419
Royalties	0
Land	3,276
Working capital	5,830
Total capital investment	192,105
Dollars per kW of generation capacity	192.11

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 2 spare absorber trains, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE A-16. LIMESTONE PROCESS, CASE 8,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1000-MW, 3.5% sulfur, 1979 NSPS, spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	276.7 ktons	8.50/ton	2,352
Total raw materials cost			2,352
Conversion costs			
Operating labor and supervision			
FGD	41,950 man-hr	15.00/man-hr	629
Utilities			
Steam	1,051,000 klb	2.50/klb	2,627
Process water	515,700 kgal	0.14/kgal	72
Electricity	95,875,000 kWh	0.037/kWh	3,548
Maintenance			
Labor and material			6,782
Analysis	6,590 man-hr	21.00/man-hr	138
Total conversion costs			13,796
Total direct costs			16,148
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			4,530
Total first-year operating and maintenance costs			20,678
Levelized capital charges (14.7% of total capital investment)			28,239
Total first-year annual revenue requirements			48,917
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			38,998
Levelized capital charges (14.7% of total capital investment)			28,239
Levelized annual revenue requirements			67,237

	M$	Mills/kWh
First-year annual revenue requirements	48.92	8.89
Levelized annual revenue requirements	67.24	12.23

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

APPENDIX B

LIMESTONE PROCESS WITH FORCED OXIDATION

TABLE B-1. LIMESTONE PROCESS, CASE 9,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(200-MW, 0.7% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	981
Feed preparation	1,589
Gas handling	6,489
SO_2 absorption	10,353
Oxidation	0
Reheat	0
Solids separation	1,790
Total process capital	21,202
Services, utilities, and miscellaneous	1,272
Total direct investment excluding landfill	22,474
Landfill construction	661
Landfill equipment	379
Total direct investment	23,514
Indirect Investment	
Engineering design and supervision	1,798
Architect and engineering contractor	674
Construction expense	4,045
Contractor fees	1,348
Contingency	3,034
Waste disposal indirect investment	351
Total fixed investment	34,764
Other Capital Investment	
Allowance for startup and modifications	2,671
Interest during construction	5,423
Royalties	0
Land	114
Working capital	1,419
Total capital investment	44,391
Dollars per kW of generation capacity	221.96

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.

Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.

Reheat to 175°F as required.

Capital costs begin with an inlet plenum downstream of the ESP.

Stack plenum and stack excluded.

Disposal site one mile from power plant.

TABLE B-2. LIMESTONE PROCESS, CASE 9, WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, 0.7% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	8.9 ktons	8.50/ton	76
Total raw materials cost			76
Conversion costs			
Operating labor and supervision			
FGD	30,660 man-hr	15.00/man-hr	460
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Process water	55,180 kgal	0.14/kgal	8
Electricity	16,825,000 kWh	0.037/kWh	622
Diesel fuel	35,360 gal	1.60/gal	57
Maintenance			
Labor and material			2,054
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			3,794
Total direct costs			3,870
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,864
Total first-year operating and maintenance costs			5,734
Levelized capital charges (14.7% of total capital investment)			6,526
Total first-year annual revenue requirements			12,260
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			10,815
Levelized capital charges (14.7% of total capital investment)			6,526
Levelized annual revenue requirements			17,341

	M$	Mills/kWh
First-year annual revenue requirements	12.26	11.15
Levelized annual revenue requirements	17.34	15.76

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-3. LIMESTONE PROCESS, CASE 10,
WITH FORCED OXIDATION
SUMMARY OF CAPITAL INVESTMENT
(200-MW, ~~3~~2.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,301
Feed preparation	2,464
Gas handling	5,745
SO_2 absorption	9,302
Oxidation	1,185
Reheat	1,921
Solids separation	2,382
Total process capital	24,300
Services, utilities, and miscellaneous	1,458
Total direct investment excluding landfill	25,758
Landfill construction	1,377
Landfill equipment	453
Total direct investment	27,588
Indirect Investment	
Engineering design and supervision	2,061
Architect and engineering contractor	773
Construction expense	4,636
Contractor fees	1,545
Contingency	3,477
Waste disposal indirect investment	664
Total fixed investment	40,744
Other Capital Investment	
Allowance for startup and modifications	3,060
Interest during construction	6,356
Royalties	0
Land	228
Working capital	1,702
Total capital investment	52,090
Dollars per kW of generation capacity	260.45

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-4. LIMESTONE PROCESS, CASE 10,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, 2.0% sulfur, 1979 NSPS, spray tower,
landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	30.5 ktons	8.50/ton	259
Total raw materials cost			259
Conversion costs			
Operating labor and supervision			
FGD	33,470 man-hr	15.00/man-hr	502
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	225,000 klb	2.50/klb	562
Process water	72,000 kgal	0.14/kgal	10
Electricity	17,259,000 kWh	0.037/kWh	639
Diesel fuel	36,930 gal	1.60/gal	59
Maintenance			
Labor and material			2,374
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			4,739
Total direct costs			4,998
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,082
Total first-year operating and maintenance costs			7,080
Levelized capital charges (14.7% of total capital investment)			7,657
Total first-year annual revenue requirements			14,737
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			13,353
Levelized capital charges (14.7% of total capital investment)			7,657
Levelized annual revenue requirements			21,010

	M$	Mills/kWh
First-year annual revenue requirements	14.74	13.40
Levelized annual revenue requirements	21.01	19.10

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-5. LIMESTONE PROCESS, CASE 11,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(200-MW, 3.5% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,282
Feed preparation	2,746
Gas handling	5,980
SO_2 absorption	10,778
Oxidation	1,523
Reheat	1,934
Solids separation	2,780
Total process capital	27,023
Services, utilities, and miscellaneous	1,622
Total direct investment excluding landfill	28,645
Landfill construction	2,097
Landfill equipment	569
Total direct investment	31,311
Indirect Investment	
Engineering design and supervision	2,292
Architect and engineering contractor	859
Construction expense	5,156
Contractor fees	1,719
Contingency	3,867
Waste disposal indirect investment	986
Total fixed investment	46,190
Other Capital Investment	
Allowance for startup and modifications	3,403
Interest during construction	7,206
Royalties	0
Land	344
Working capital	1,924
Total capital investment	59,067
Dollars per kW of generation capacity	295.34

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-6. LIMESTONE PROCESS, CASE 11,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, 3.5% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	58.3 ktons	8.50/ton	496
Total raw materials cost			496
Conversion costs			
Operating labor and supervision			
FGD	33,470 man-hr	15.00/man-hr	502
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	221,400 klb	2.50/klb	554
Process water	79,010 kgal	0.14/kgal	11
Electricity	24,427,000 kWh	0.037/kWh	904
Diesel fuel	47,340 gal	1.20/gal	76
Maintenance			
Labor and material			2,658
Analysis	3,310 man-hr	21.00/man-hr	69
Total conversion costs			5,298
Total direct costs			5,794
Indirect Costs - First Year			
Overheads			
Plant and administrative			2,252
Total first-year operating and maintenance costs			8,046
Levelized capital charges (14.7% of total capital investment)			8,683
Total first-year annual revenue requirements			16,729
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			15,175
Levelized capital charges (14.7% of total capital investment)			8,683
Levelized annual revenue requirements			23,858

	M$	Mills/kWh
First-year annual revenue requirements	16.73	15.21
Levelized annual revenue requirements	23.86	21.69

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-7. LIMESTONE PROCESS, CASE 12,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(200-MW, 5.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,370
Feed preparation	3,410
Gas handling	5,993
SO_2 absorption	10,880
Oxidation	1,440
Reheat	1,919
Solids separation	3,105
Total process capital	28,117
Services, utilities, and miscellaneous	1,687
Total direct investment excluding landfill	29,804
Landfill construction	2,655
Landfill equipment	767
Total direct investment	33,226
Indirect Investment	
Engineering design and supervision	2,384
Architect and engineering contractor	894
Construction expense	5,365
Contractor fees	1,788
Contingency	4,024
Waste disposal indirect investment	1,258
Total fixed investment	48,939
Other Capital Investment	
Allowance for startup and modifications	3,541
Interest during construction	7,634
Royalties	0
Land	432
Working capital	2,040
Total capital investment	62,586
Dollars per kW of generation capacity	312.93

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-8. LIMESTONE PROCESS, CASE 12,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, 5.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	83.4 ktons	8.50/ton	709
Total raw materials cost			709
Conversion costs			
Operating labor and supervision			
FGD	33,470 man-hr	15.00/man-hr	502
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	217,900 klb	2.50/klb	545
Process water	85,120 kgal	0.14/kgal	12
Electricity	26,946,000 kWh	0.037/kWh	997
Diesel fuel	69,480 gal	1.60/gal	111
Maintenance			
Labor and material			2,785
Analysis	3,320 man-hr	21.00/man-hr	70
Total conversion costs			5,546
Total direct costs			6,255
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,329
Total first-year operating and maintenance costs			8,584
Levelized capital charges (14.7% of total capital investment)			9,200
Total first-year annual revenue requirements			17,784
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			16,188
Levelized capital charges (14.7% of total capital investment)			9,200
Levelized annual revenue requirements			25,388

	M$	Mills/kWh
First-year annual revenue requirements	17.78	16.17
Levelized annual revenue requirements	25.39	23.08

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-9. LIMESTONE PROCESS, CASE 13,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 0.7% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,265
Feed preparation	2,364
Gas handling	11,118
SO_2 absorption	17,992
Oxidation	0
Reheat	0
Solids separation	2,263
Total process capital	35,002
Services, utilities, and miscellaneous	2,100
Total direct investment excluding landfill	37,102
Landfill construction	1,124
Landfill equipment	411
Total direct investment	38,637
Indirect Investment	
Engineering design and supervision	2,597
Architect and engineering contractor	742
Construction expense	5,936
Contractor fees	1,855
Contingency	4,823
Waste disposal indirect investment	523
Total fixed investment	55,113
Other Capital Investment	
Allowance for startup and modifications	4,244
Interest during construction	8,598
Royalties	0
Land	190
Working capital	2,208
Total capital investment	70,353
Dollars per kW of generation capacity	140.71

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-10. LIMESTONE PROCESS, CASE 13,
WITH FORCED OXIDATION
SUMMARY OF ANNUAL REVENUE REQUIREMENTS
(500-MW, 0.7% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	21.9 ktons	8.50/ton	186
Total raw materials cost			186
Conversion costs			
Operating labor and supervision			
FGD	39,620 man-hr	15.00/man-hr	594
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Process water	135,400 kgal	0.14/kgal	19
Electricity	39,723,000 kWh	0.037/kWh	1,470
Diesel fuel	36,310 gal	1.60/gal	58
Maintenance			
Labor and material			3,015
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			5,749
Total direct costs			5,935
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,521
Total first-year operating and maintenance costs			8,456
Levelized capital charges (14.7% of total capital investment)			10,342
Total first-year annual revenue requirements			18,798
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			15,948
Levelized capital charges (14.7% of total capital investment)			10,342
Levelized annual revenue requirements			26,290

	M$	Mills/kWh
First-year annual revenue requirements	18.80	6.84
Levelized annual revenue requirements	26.29	9.56

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-11. LIMESTONE PROCESS, CASE 14, WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 2.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,310
Feed preparation	3,157
Gas handling	10,831
SO_2 absorption	17,782
Oxidation	2,238
Reheat	3,615
Solids separation	2,990
Total process capital	41,923
Services, utilities, and miscellaneous	2,515
Total direct investment excluding landfill	44,438
Landfill construction	2,453
Landfill equipment	606
Total direct investment	47,497
Indirect Investment	
Engineering design and supervision	3,111
Architect and engineering contractor	889
Construction expense	7,110
Contractor fees	2,220
Contingency	5,777
Waste disposal indirect investment	1,084
Total fixed investment	67,688
Other Capital Investment	
Allowance for startup and modifications	5,084
Interest during construction	10,559
Royalties	0
Land	402
Working capital	2,833
Total capital investment	86,566
Dollars per kW of generation capacity	173.13

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-12. LIMESTONE PROCESS, CASE 14,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 2.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	74.6 ktons	8.50/ton	634
Total raw materials cost			634
Conversion costs			
Operating labor and supervision			
FGD	43,870 man-hr	15.00/man-hr	658
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	550,900 klb	2.50/klb	1,377
Process water	176,300 kgal	0.14/kgal	25
Electricity	39,536,000 kWh	0.037/kWh	1,463
Diesel fuel	48,550 gal	1.60/gal	78
Maintenance			
Labor and material			3,647
Analysis	4,970 man-hr	21.00/man-hr	104
Total conversion costs			7,876
Total direct costs			8,510
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,961
Total first-year operating and maintenance costs			11,471
Levelized capital charges (14.7% of total capital investment)			12,725
Total first-year annual revenue requirements			24,196
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			21,636
Levelized capital charges (14.7% of total capital investment)			12,725
Levelized annual revenue requirements			34,361

	M$	Mills/kWh
First-year annual revenue requirements	24.20	8.80
Levelized annual revenue requirements	34.36	12.49

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-13. LIMESTONE PROCESS, CASE 15,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower,
landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,528
Feed preparation	4,715
Gas handling	11,281
SO_2 absorption	20,288
Oxidation	2,670
Reheat	3,634
Solids separation	3,679
Total process capital	48,795
Services, utilities, and miscellaneous	2,928
Total direct investment excluding landfill	51,723
Landfill construction	3,781
Landfill equipment	1,123
Total direct investment	56,627
Indirect Investment	
Engineering design and supervision	3,621
Architect and engineering contractor	1,034
Construction expense	8,276
Contractor fees	2,586
Contingency	6,724
Waste disposal indirect investment	1,707
Total fixed investment	80,575
Other Capital Investment	
Allowance for startup and modifications	5,917
Interest during construction	12,570
Royalties	0
Land	614
Working capital	3,392
Total capital investment	103,068
Dollars per kW of generation capacity	206.14

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-14. LIMESTONE PROCESS, CASE 15,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower,
landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	142.9 ktons	8.50/ton	1,214
Total raw materials cost			1,214
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	542,200 klb	2.50/klb	1,356
Process water	193,400 kgal	0.14/kgal	27
Electricity	57,657,000 kWh	0.037/kWh	2,133
Diesel fuel	103,200 gal	1.60/gal	165
Maintenance			
Labor and material			4,285
Analysis	4,990 man-hr	21.00/man-hr	105
Total conversion costs			9,340
Total direct costs			10,554
Indirect Costs - First Year			
Overheads			
Plant and administrative			3,395
Total first-year operating and maintenance costs			13,949
Levelized capital charges (14.7% of total capital investment)			15,151
Total first-year annual revenue requirements			29,100
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			26,309
Levelized capital charges (14.7% of total capital investment)			15,151
Levelized annual revenue requirements			41,460

	M$	Mills/kWh
First-year annual revenue requirements	29.10	10.58
Levelized annual revenue requirements	41.46	15.08

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-15. LIMESTONE PROCESS, CASE 16, WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 5.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,825
Feed preparation	5,326
Gas handling	11,285
SO_2 absorption	20,483
Oxidation	3,039
Reheat	3,604
Solids separation	4,192
Total process capital	50,754
Services, utilities, and miscellaneous	3,045
Total direct investment excluding landfill	53,799
Landfill construction	4,807
Landfill equipment	1,401
Total direct investment	60,007
Indirect Investment	
Engineering design and supervision	3,766
Architect and engineering contractor	1,076
Construction expense	8,607
Contractor fees	2,690
Contingency	6,994
Waste disposal indirect investment	2,165
Total fixed investment	85,305
Other Capital Investment	
Allowance for startup and modifications	6,155
Interest during construction	13,308
Royalties	0
Land	778
Working capital	3,625
Total capital investment	109,171
Dollars per kW of generation capacity	218.34

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-16. LIMESTONE PROCESS, CASE 16,
WITH FORCED OXIDATION
SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 5.0% sulfur, 1979 NSPS, spray tower,
landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	204.3 ktons	8.50/ton	1,736
Total raw materials cost			1,736
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	659
Waste disposal	33,280 man-hr	21.00/man-hr	699
Utilities			
Steam	533,600 klb	2.50/klb	1,334
Process water	208,400 kgal	0.14/kgal	29
Electricity	63,969,000 kWh	0.037/kWh	2,367
Diesel fuel	120,900 gal	1.60/gal	193
Maintenance			
Labor and material			4,490
Analysis	5,000 man-hr	21.00/man-hr	105
Total conversion costs			9,876
Total direct costs			11,612
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,571
Total first-year operating and maintenance costs			15,183
Levelized capital charges (14.7% of total capital investment)			16,048
Total first-year annual revenue requirements			31,231
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			28,635
Levelized capital charges (14.7% of total capital investment)			16,048
Levelized annual revenue requirements			44,683

	M$	Mills/kWh
First-year annual revenue requirements	31.23	11.36
Levelized annual revenue requirements	44.68	16.25

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-17. LIMESTONE PROCESS, CASE 17,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, TCA, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,531
Feed preparation	4,754
Gas handling	11,489
SO_2 absorption	18,266
Oxidation	2,171
Reheat	3,526
Solids separation	3,663
Total process capital	46,400
Services, utilities, and miscellaneous	2,784
Total direct investment excluding landfill	49,184
Landfill construction	3,818
Landfill equipment	1,128
Total direct investment	54,130
Indirect Investment	
Engineering design and supervision	3,443
Architect and engineering contractor	984
Construction expense	7,870
Contractor fees	2,459
Contingency	6,394
Waste disposal indirect investment	1,722
Total fixed investment	77,002
Other Capital Investment	
Allowance for startup and modifications	5,627
Interest during construction	12,012
Royalties	0
Land	620
Working capital	3,230
Total capital investment	98,491
Dollars per kW of generation capacity	196.98

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-18. LIMESTONE PROCESS, CASE 17,
WITH FORCED OXIDATION
SUMMARY OF ANNUAL REVENUE REQUIREMENTS
(500-MW, 3.5% sulfur, 1979 NSPS, TCA, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	145.9 ktons	8.50/ton	1,240
Total raw materials cost			1,240
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	526,500 klb	2.50/klb	1,316
Process water	193,900 kgal	0.14/kgal	27
Electricity	48,405,000 kWh	0.037/kWh	1,791
Diesel fuel	103,400 gal	1.60/gal	165
Maintenance			
Labor and material			4,084
Analysis	4,990 man-hr	21.00/man-hr	105
Total conversion costs			8,757
Total direct costs			9,997
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,274
Total first-year operating and maintenance costs			13,271
Levelized capital charges (14.7% of total capital investment)			14,478
Total first-year annual revenue requirements			27,749
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			25,030
Levelized capital charges (14.7% of total capital investment)			14,478
Levelized annual revenue requirements			39,508

	M$	Mills/kWh
First-year annual revenue requirements	27.75	10.09
Levelized annual revenue requirements	39.51	14.37

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-19. LIMESTONE PROCESS, CASE 18,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,492
Feed preparation	4,643
Gas handling	13,358
SO_2 absorption	23,235
Oxidation	1,354
Reheat	3,534
Solids separation	3,599
Total process capital	52,215
Services, utilities, and miscellaneous	3,133
Total direct investment excluding landfill	55,348
Landfill construction	3,657
Landfill equipment	924
Total direct investment	59,929
Indirect Investment	
Engineering design and supervision	3,874
Architect and engineering contractor	1,107
Construction expense	8,856
Contractor fees	2,767
Contingency	7,195
Waste disposal indirect investment	1,618
Total fixed investment	85,346
Other Capital Investment	
Allowance for startup and modifications	6,332
Interest during construction	13,314
Royalties	0
Land	594
Working capital	3,595
Total capital investment	109,181
Dollars per kW of generation capacity	218.36

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-20. LIMESTONE PROCESS, CASE 18,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	132.6 ktons	8.50/ton	1,128
Total raw materials cost			1,128
Conversion costs			
Operating labor and supervision			
FGD	45,510 man-hr	15.00/man-hr	683
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	492,700 klb	2.50/klb	1,232
Process water	192,000 kgal	0.14/kgal	27
Electricity	73,142,000 kWh	0.037/kWh	2,706
Diesel fuel	77,030 gal	1.60/gal	123
Maintenance			
Labor and material			4,565
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			10,052
Total direct costs			11,180
Indirect Costs - First Year			
Overheads			
Plant and administrative			3,578
Total first-year operating and maintenance costs			14,758
Levelized capital charges (14.7% of total capital investment)			16,050
Total first-year annual revenue requirements			30,808
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			27,834
Levelized capital charges (14.7% of total capital investment)			16,050
Levelized annual revenue requirements			43,884

	M$	Mills/kWh
First-year annual revenue requirements	30.81	11.20
Levelized annual revenue requirements	43.88	15.96

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-21. LIMESTONE PROCESS, CASE 19,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, no reheat, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,528
Feed preparation	4,715
Gas handling	11,118
SO_2 absorption	19,686
Oxidation	2,676
Reheat	0
Solids separation	3,679
Total process capital	44,402
Services, utilities, and miscellaneous	2,664
Total direct investment excluding landfill	47,066
Landfill construction	3,782
Landfill equipment	1,123
Total direct investment	51,971
Indirect Investment	
Engineering design and supervision	3,295
Architect and engineering contractor	941
Construction expense	7,531
Contractor fees	2,353
Contingency	6,119
Waste disposal indirect investment	1,706
Total fixed investment	73,916
Other Capital Investment	
Allowance for startup and modifications	5,384
Interest during construction	11,531
Royalties	0
Land	614
Working capital	3,003
Total capital investment	94,448
Dollars per kW of generation capacity	188.90

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-22. LIMESTONE PROCESS, CASE 19,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, no reheat, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	142.9 ktons	8.50/ton	1,214
Total raw materials cost			1,214
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Process water	193,400 kgal	0.14/kgal	27
Electricity	56,525,000 kWh	0.037/kWh	2,091
Diesel fuel	103,200 gal	1.60/gal	165
Maintenance			
Labor and material			3,913
Analysis	4,990 man-hr	21.00/man-hr	105
Total conversion costs			7,570
Total direct costs			8,784
Indirect Costs - First Year			
Overheads			
Plant and administrative			3,172
Total first-year operating and maintenance costs			11,956
Levelized capital charges (14.7% of total capital investment)			13,884
Total first-year annual revenue requirements			25,840
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			22,550
Levelized capital charges (14.7% of total capital investment)			13,884
Levelized annual revenue requirements			36,434

	M$	Mills/kWh
First-year annual revenue requirements	25.84	9.40
Levelized annual revenue requirements	36.43	13.25

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-23. LIMESTONE PROCESS, CASE 20,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(1000-MW, 0.7% sulfur, 1979 NSPS, spray tower,
landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,343
Feed preparation	2,796
Gas handling	19,038
SO_2 absorption	30,949
Oxidation	0
Reheat	0
Solids separation	2,580
Total process capital	56,706
Services, utilities, and miscellaneous	3,402
Total direct investment excluding landfill	60,108
Landfill construction	1,708
Landfill equipment	503
Total direct investment	62,319
Indirect Investment	
Engineering design and supervision	3,607
Architect and engineering contractor	601
Construction expense	8,415
Contractor fees	2,404
Contingency	7,514
Waste disposal indirect investment	729
Total fixed investment	85,589
Other Capital Investment	
Allowance for startup and modifications	6,611
Interest during construction	13,352
Royalties	0
Land	285
Working capital	3,390
Total capital investment	109,227
Dollars per kW of generation capacity	109.23

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 2 spare absorber trains, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-24. LIMESTONE PROCESS, CASE 20,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1000-MW, 0.7% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	42.5 ktons	8.50/ton	361
Total raw materials cost			361
Conversion costs			
Operating labor and supervision			
FGD	49,330 man-hr	15.00/man-hr	740
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Process water	263,000 kgal	0.14/kgal	37
Electricity	75,394,000 kWh	0.037/kWh	2,790
Diesel fuel	41,530 gal	1.60/gal	66
Maintenance			
Labor and material			4,274
Analysis	4,960 man-hr	21.00/man-hr	104
Total conversion costs			8,535
Total direct costs			8,896
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,385
Total first-year operating and maintenance costs			12,281
Levelized capital charges (14.7% of total capital investment)			16,056
Total first-year annual revenue requirements			28,337
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			23,163
Levelized capital charges (14.7% of total capital investment)			16,056
Levelized annual revenue requirements			39,219

	M$	Mills/kWh
First-year annual revenue requirements	28.34	5.15
Levelized annual revenue requirements	39.22	7.13

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-25. LIMESTONE PROCESS, CASE 21,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(1000-MW, 2.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,544
Feed preparation	4,890
Gas handling	20,564
SO_2 absorption	33,920
Oxidation	4,437
Reheat	6,906
Solids separation	3,666
Total process capital	76,927
Services, utilities, and miscellaneous	4,616
Total direct investment excluding landfill	81,543
Landfill construction	3,795
Landfill equipment	1,124
Total direct investment	86,462
Indirect Investment	
Engineering design and supervision	4,893
Architect and engineering contractor	815
Construction expense	11,416
Contractor fees	3,262
Contingency	10,193
Waste disposal indirect investment	1,621
Total fixed investment	118,662
Other Capital Investment	
Allowance for startup and modifications	8,970
Interest during construction	18,511
Royalties	0
Land	621
Working capital	4,892
Total capital investment	151,656
Dollars per kW of generation capacity	151.66

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 2 spare absorber trains, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-26. LIMESTONE PROCESS, CASE 21,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1000-MW, 2.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	144.9 ktons	8.50/ton	1,232
Total raw materials cost			1,232
Conversion costs			
Operating labor and supervision			
FGD	55,130 man-hr	15.00/man-hr	827
Waste disposal	29,120 man-hr	21.00/man-hr	612
Utilities			
Steam	1,067,000 klb	2.50/klb	2,668
Process water	341,500 kgal	0.14/kgal	48
Electricity	75,486,000 kWh	0.037/kWh	2,793
Diesel fuel	103,200 gal	1.60/gal	165
Maintenance			
Labor and material			5,855
Analysis	6,630 man-hr	21.00/man-hr	139
Total conversion costs			13,107
Total direct costs			14,339
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			4,460
Total first-year operating and maintenance costs			18,799
Levelized capital charges (14.7% of total capital investment)			22,294
Total first-year annual revenue requirements			41,093
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			35,455
Levelized capital charges (14.7% of total capital investment)			22,294
Levelized annual revenue requirements			57,749

	M$	Mills/kWh
First-year annual revenue requirements	41.09	7.47
Levelized annual revenue requirements	57.75	10.50

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-27. LIMESTONE PROCESS, CASE 22,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(1000-MW, 3.5% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,915
Feed preparation	6,039
Gas handling	21,738
SO_2 absorption	39,913
Oxidation	5,467
Reheat	7,113
Solids separation	4,714
Total process capital	87,899
Services, utilities, and miscellaneous	5,274
Total direct investment excluding landfill	93,173
Landfill construction	5,877
Landfill equipment	2,009
Total direct investment	101,059
Indirect Investment	
Engineering design and supervision	5,590
Architect and engineering contractor	932
Construction expense	13,044
Contractor fees	3,727
Contingency	11,647
Waste disposal indirect investment	2,565
Total fixed investment	138,564
Other Capital Investment	
Allowance for startup and modifications	10,249
Interest during construction	21,616
Royalties	0
Land	951
Working capital	5,840
Total capital investment	177,220
Dollars per kW of generation capacity	177.22

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 2 spare absorber trains, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-28. LIMESTONE PROCESS, CASE 22,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1000-MW, 3.5% sulfur, 1979 NSPS, spray tower,
landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	276.7 ktons	8.50/ton	2,352
Total raw materials cost			2,352
Conversion costs			
Operating labor and supervision			
FGD	55,130 man-hr	15.00/man-hr	827
Waste disposal	45,760 man-hr	21.00/man-hr	961
Utilities			
Steam	1,050,000 klb	2.50/klb	2,625
Process water	374,700 kgal	0.14/kgal	52
Electricity	109,243,000 kWh	0.037/kWh	4,042
Diesel fuel	183,500 gal	1.60/gal	294
Maintenance			
Labor and material			6,759
Analysis	6,670 man-hr	21.00/man-hr	140
Total conversion costs			15,700
Total direct costs			18,052
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			5,212
Total first-year operating and maintenance costs			23,264
Levelized capital charges (14.7% of total capital investment)			26,051
Total first-year annual revenue requirements			49,315
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			43,876
Levelized capital charges (14.7% of total capital investment)			26,051
Levelized annual revenue requirements			69,927

	M$	Mills/kWh
First-year annual revenue requirements	49.32	8.97
Levelized annual revenue requirements	69.93	12.71

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE B-29. LIMESTONE PROCESS, CASE 23, WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(1000-MW, 5.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	3,260
Feed preparation	6,892
Gas handling	21,811
SO_2 absorption	40,341
Oxidation	6,138
Reheat	7,055
Solids separation	5,839
Total process capital	91,336
Services, utilities, and miscellaneous	5,480
Total direct investment excluding landfill	96,816
Landfill construction	7,483
Landfill equipment	2,265
Total direct investment	106,564
Indirect Investment	
Engineering design and supervision	5,809
Architect and engineering contractor	968
Construction expense	13,554
Contractor fees	3,873
Contingency	12,102
Waste disposal indirect investment	3,207
Total fixed investment	146,077
Other Capital Investment	
Allowance for startup and modifications	10,650
Interest during construction	22,788
Royalties	0
Land	1,205
Working capital	6,240
Total capital investment	186,960
Dollars per kW of generation capacity	186.96

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 2 spare absorber trains, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE B-30. LIMESTONE PROCESS, CASE 23,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1000-MW, 5.0% sulfur, 1979 NSPS, spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	395.6 ktons	8.50/ton	3,363
Total raw materials cost			3,363
Conversion costs			
Operating labor and supervision			
FGD	55,130 man-hr	15.00/man-hr	827
Waste disposal	54,080 man-hr	21.00/man-hr	1,136
Utilities			
Steam	1,033,000 klb	2.50/klb	2,584
Process water	403,700 kgal	0.14/kgal	56
Electricity	121,399,000 kWh	0.037/kWh	4,492
Diesel fuel	188,900 gal	1.60/gal	302
Maintenance			
Labor and material			7,069
Analysis	6,710 man-hr	21.00/man-hr	141
Total conversion costs			16,607
Total direct costs			19,970
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			5,504
Total first-year operating and maintenance costs			25,474
Levelized capital charges (14.7% of total capital investment)			27,483
Total first-year annual revenue requirements			52,957
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			48,043
Levelized capital charges (14.7% of total capital investment)			27,483
Levelized annual revenue requirements			75,526

	M$	Mills/kWh
First-year annual revenue requirements	52.96	9.63
Levelized annual revenue requirements	75.53	13.73

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

APPENDIX C

LIMESTONE PROCESS WITH FORCED OXIDATION AND ADDITIVE

TABLE C-1. LIMESTONE PROCESS, CASE 24, WITH FORCED OXIDATION AND ADDITIVE SUMMARY OF CAPITAL INVESTMENT

(200-MW, 3.5% sulfur, 1979 NSPS, spray tower, adipic acid, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,382
Feed preparation	2,788
Gas handling	5,558
SO_2 absorption	9,455
Oxidation	1,351
Reheat	1,724
Solids separation	2,682
Total process capital	24,940
Services, utilities, and miscellaneous	1,497
Total direct investment excluding landfill	26,437
Landfill construction	1,870
Landfill equipment	516
Total direct investment	28,823
Indirect Investment	
Engineering design and supervision	2,115
Architect and engineering contractor	793
Construction expense	4,759
Contractor fees	1,586
Contingency	3,569
Waste disposal indirect investment	881
Total fixed investment	42,526
Other Capital Investment	
Allowance for startup and modifications	3,140
Interest during construction	6,634
Royalties	0
Land	307
Working capital	1,764
Total capital investment	54,371
Dollars per kW of generation capacity	271.85

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-2. LIMESTONE PROCESS, CASE 24,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(200-MW, 3.5% sulfur, 1979 NSPS, spray tower,
adipic acid, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	44.5 ktons	8.50/ton	379
Adipic acid	56.7 tons	1,500.00/ton	85
Total raw materials cost			464
Conversion costs			
Operating labor and supervision			
FGD	32,840 man-hr	15.00/man-hr	493
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	173,600 klb	2.50/klb	434
Process water	72,800 kgal	0.14/kgal	10
Electricity	19,260,000 kWh	0.037/kWh	713
Diesel fuel	41,890 gal	1.60/gal	67
Maintenance			
Labor and material			2,451
Analysis	3,310 man-hr	21.00/man-hr	69
Total conversion costs			4,761
Total direct costs			5,225
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,122
Total first-year operating and maintenance costs			7,347
Levelized capital charges (14.7% of total capital investment)			7,993
Total first-year annual revenue requirements			15,340
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			13,856
Levelized capital charges (14.7% of total capital investment)			7,993
Levelized annual revenue requirements			21,849

	M$	Mills/kWh
First-year annual revenue requirements	15.34	13.95
Levelized annual revenue requirements	21.85	19.86

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-3. LIMESTONE PROCESS, CASE 25,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 0.7% sulfur, 1979 NSPS, spray tower,
adipic acid, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,045
Feed preparation	1,960
Gas handling	10,372
SO_2 absorption	15,152
Oxidation	0
Reheat	0
Solids separation	1,954
Total process capital	30,483
Services, utilities, and miscellaneous	1,829
Total direct investment excluding landfill	32,312
Landfill construction	1,008
Landfill equipment	404
Total direct investment	33,724
Indirect Investment	
Engineering design and supervision	2,262
Architect and engineering contractor	646
Construction expense	5,170
Contractor fees	1,616
Contingency	4,201
Waste disposal indirect investment	476
Total fixed investment	48,095
Other Capital Investment	
Allowance for startup and modifications	3,696
Interest during construction	7,503
Royalties	0
Land	171
Working capital	1,929
Total capital investment	61,394
Dollars per kW of generation capacity	122.79

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-4. LIMESTONE PROCESS, CASE 25,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 0.7% sulfur, 1979 NSPS, spray tower,
adipic acid, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	16.7 ktons	8.50/ton	142
Adipic acid	21.2 tons	1,500.00/ton	32
Total raw materials cost			174
Conversion costs			
Operating labor and supervision			
FGD	38,980 man-hr	15.00/man-hr	585
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Process water	127,700 kgal	0.14/kgal	18
Electricity	28,653,000 kWh	0.037/kWh	1,060
Diesel fuel	36,130 gal	1.60/gal	58
Maintenance			
Labor and material			2,627
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			4,941
Total direct costs			5,115
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,283
Total first-year operating and maintenance costs			7,398
Levelized capital charges (14.7% of total capital investment)			9,025
Total first-year annual revenue requirements			16,423
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			13,953
Levelized capital charges (14.7% of total capital investment)			9,025
Levelized annual revenue requirements			22,978

	M$	Mills/kWh
First-year annual revenue requirements	16.42	5.97
Levelized annual revenue requirements	22.98	8.36

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-5. LIMESTONE PROCESS, CASE 26,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 2.0% sulfur, 1979 NSPS, spray tower,
adipic acid, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,337
Feed preparation	2,838
Gas handling	9,808
SO_2 absorption	16,607
Oxidation	2,344
Reheat	2,593
Solids separation	2,862
Total process capital	38,389
Services, utilities, and miscellaneous	2,303
Total direct investment excluding landfill	40,692
Landfill construction	2,178
Landfill equipment	576
Total direct investment	43,446
Indirect Investment	
Engineering design and supervision	2,848
Architect and engineering contractor	814
Construction expense	6,511
Contractor fees	2,035
Contingency	5,290
Waste disposal indirect investment	969
Total fixed investment	61,913
Other Capital Investment	
Allowance for startup and modifications	4,654
Interest during construction	9,658
Royalties	0
Land	358
Working capital	2,526
Total capital investment	79,109
Dollars per kW of generation capacity	158.22

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-6. LIMESTONE PROCESS, CASE 26,
WITH FORCED OXIDATION AND ADDITIVE
SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 2.0% sulfur, 1979 NSPS, spray tower, adipic acid, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	56.9 ktons	8.50/ton	484
Adipic acid	71.8 tons	1,500.00/ton	108
Total raw materials cost			592
Conversion costs			
Operating labor and supervision			
FGD	41,590 man-hr	15.00/man-hr	624
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	267,800 klb	2.50/klb	669
Process water	149,100 kgal	1.60/kgal	21
Electricity	35,916,000 kWh	0.037/kWh	1,329
Diesel fuel	47,550 gal	1.60/gal	76
Maintenance			
Labor and material			3,338
Analysis	3,310 man-hr	21.00/man-hr	70
Total conversion costs			6,651
Total direct costs			7,243
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,733
Total first-year operating and maintenance costs			9,976
Levelized capital charges (14.7% of total capital investment)			11,629
Total first-year annual revenue requirements			21,605
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			18,814
Levelized capital charges (14.7% of total capital investment)			11,629
Levelized annual revenue requirements			30,443

	M$	Mills/kWh
First-year annual revenue requirements	21.61	7.86
Levelized annual revenue requirements	30.44	11.07

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-7. LIMESTONE PROCESS, CASE 27,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, adipic acid, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,552
Feed preparation	4,296
Gas handling	10,476
SO_2 absorption	17,726
Oxidation	2,348
Reheat	3,221
Solids separation	3,456
Total process capital	44,075
Services, utilities, and miscellaneous	2,644
Total direct investment excluding landfill	46,719
Landfill construction	3,362
Landfill equipment	890
Total direct investment	50,971
Indirect Investment	
Engineering design and supervision	3,270
Architect and engineering contractor	934
Construction expense	7,475
Contractor fees	2,336
Contingency	6,073
Waste disposal indirect investment	1,496
Total fixed investment	72,555
Other Capital Investment	
Allowance for startup and modifications	5,345
Interest during construction	11,319
Royalties	0
Land	548
Working capital	3,003
Total capital investment	92,770
Dollars per kW of generation capacity	185.54

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-8. LIMESTONE PROCESS, CASE 27,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, adipic acid, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	109.1 ktons	8.50/ton	927
Adipic acid	138.9 tons	1,500.00/ton	208
Total raw materials cost			1,135
Conversion costs			
Operating labor and supervision			
FGD	42,920 man-hr	15.00/man-hr	644
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	425,200 klb	2.50/klb	1,063
Process water	178,300 kgal	0.14/kgal	25
Electricity	45,024,000 kWh	0.037/kWh	1,666
Diesel fuel	73,150 gal	1.60/gal	117
Maintenance			
Labor and material			3,865
Analysis	3,330 man-hr	21.00/man-hr	70
Total conversion costs			7,974
Total direct costs			9,109
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,062
Total first-year operating and maintenance costs			12,171
Levelized capital charges (14.7% of total capital investment)			13,637
Total first-year annual revenue requirements			25,808
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			22,954
Levelized capital charges (14.7% of total capital investment)			13,637
Levelized annual revenue requirements			36,591

	M$	Mills/kWh
First-year annual revenue requirements	25.81	9.38
Levelized annual revenue requirements	36.59	13.31

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-9. LIMESTONE PROCESS, CASE 28, WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 5.0% sulfur, 1979 NSPS, spray tower, adipic acid, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,630
Feed preparation	4,890
Gas handling	10,510
SO_2 absorption	17,795
Oxidation	2,612
Reheat	3,224
Solids separation	3,922
Total process capital	45,583
Services, utilities, and miscellaneous	2,735
Total direct investment excluding landfill	48,318
Landfill construction	4,274
Landfill equipment	1,267
Total direct investment	53,859
Indirect Investment	
Engineering design and supervision	3,382
Architect and engineering contractor	966
Construction expense	7,731
Contractor fees	2,416
Contingency	6,281
Waste disposal indirect investment	1,929
Total fixed investment	76,564
Other Capital Investment	
Allowance for startup and modifications	5,529
Interest during construction	11,944
Royalties	0
Land	692
Working capital	3,211
Total capital investment	97,940
Dollars per kW of generation capacity	195.88

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-10. LIMESTONE PROCESS, CASE 28, WITH FORCED OXIDATION AND ADDITIVE SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 5.0% sulfur, 1979 NSPS, spray tower, adipic acid, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	156.0 ktons	8.50/ton	1,326
Adipic acid	199.3 tons	1,500.00/ton	299
Total raw materials cost			1,625
Conversion costs			
Operating labor and supervision			
FGD	42,990 man-hr	15.00/man-hr	645
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	427,000 klb	2.50/klb	1,068
Process water	192,100 kgal	0.14/kgal	27
Electricity	49,627,000 kWh	0.037/kWh	1,836
Diesel fuel	108,100 gal	1.60/gal	173
Maintenance			
Labor and material			4,032
Analysis	3,350 man-hr	21.00/man-hr	70
Total conversion costs			8,462
Total direct costs			10,087
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,215
Total first-year operating and maintenance costs			13,302
Levelized capital charges (14.7% of total capital investment)			14,397
Total first-year annual revenue requirements			27,699
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			25,088
Levelized capital charges (14.7% of total capital investment)			14,397
Levelised annual revenue requirements			39,485

	M$	Mills/kWh
First-year annual revenue requirements	27.70	10.07
Levelized annual revenue requirements	39.49	14.36

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-11. LIMESTONE PROCESS, CASE 29,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, TCA,
adipic acid, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,566
Feed preparation	4,582
Gas handling	10,447
SO_2 absorption	16,399
Oxidation	2,050
Reheat	3,138
Solids separation	3,509
Total process capital	42,691
Services, utilities, and miscellaneous	2,561
Total direct investment excluding landfill	45,252
Landfill construction	3,530
Landfill equipment	900
Total direct investment	49,682
Indirect Investment	
Engineering design and supervision	3,168
Architect and engineering contractor	905
Construction expense	7,240
Contractor fees	2,263
Contingency	5,883
Waste disposal indirect investment	1,564
Total fixed investment	70,705
Other Capital Investment	
Allowance for startup and modifications	5,177
Interest during construction	11,028
Royalties	0
Land	574
Working capital	2,946
Total capital investment	90,430
Dollars per kW of generation capacity	180.86

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-12. LIMESTONE PROCESS, CASE 29, WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, TCA, adipic acid, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	122.3 ktons	8.50/ton	1,040
Adipic acid	179.3 tons	1,500.00/ton	269
Total raw materials cost			1,309
Conversion costs			
Operating labor and supervision			
FGD	42,920 man-hr	15.00/man-hr	644
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	412,200 klb	2.50/klb	1,031
Process water	180,100 kgal	0.14/kgal	25
Electricity	39,715,000 kWh	0.037/kWh	1,470
Diesel fuel	75,910 gal	1.60/gal	121
Maintenance			
Labor and material			3,753
Analysis	3,330 man-hr	21.00/man-hr	70
Total conversion costs			7,725
Total direct costs			9,034
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,047
Total first-year operating and maintenance costs			12,081
Levelized capital charges (14.7% of total capital investment)			13,293
Total first-year annual revenue requirements			25,374
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			22,784
Levelized capital charges (14.7% of total capital investment)			13,293
Levelized annual revenue requirements			36,077

	M$	Mills/kWh
First-year annual revenue requirements	25.37	9.23
Levelized annual revenue requirements	36.08	13.12

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-13. LIMESTONE PROCESS, CASE 30, WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, adipic acid, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,562
Feed preparation	4,577
Gas handling	12,021
SO_2 absorption	18,583
Oxidation	1,386
Reheat	3,093
Solids separation	3,513
Total process capital	45,735
Services, utilities, and miscellaneous	2,744
Total direct investment excluding landfill	48,479
Landfill construction	3,529
Landfill equipment	900
Total direct investment	52,908
Indirect Investment	
Engineering design and supervision	3,394
Architect and engineering contractor	970
Construction expense	7,757
Contractor fees	2,424
Contingency	6,301
Waste disposal indirect investment	1,564
Total fixed investment	75,318
Other Capital Investment	
Allowance for startup and modifications	5,545
Interest during construction	11,750
Royalties	0
Land	573
Working capital	3,139
Total capital investment	96,325
Dollars per kW of generation capacity	192.65

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-14. LIMESTONE PROCESS, CASE 30,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, adipic acid, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	122.3 ktons	8.50/ton	1,040
Adipic acid	94.6 tons	1,500.00/ton	142
Total raw materials cost			1,182
Conversion costs			
Operating labor and supervision			
FGD	44,560 man-hr	15.00/man-hr	669
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	382,700 klb	2.50/klb	957
Process water	180,100 kgal	0.14/kgal	25
Electricity	52,571,000 kWh	0.037/kWh	1,945
Diesel fuel	75,910 gal	1.60/gal	121
Maintenance			
Labor and material			4,011
Analysis	3,330 man-hr	21.00/man-hr	70
Total conversion costs			8,409
Total direct costs			9,591
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,217
Total first-year operating and maintenance costs			12,808
Levelized capital charges (14.7% of total capital investment)			14,160
Total first-year annual revenue requirements			26,968
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			24,156
Levelized capital charges (14.7% of total capital investment)			14,160
Levelized annual revenue requirements			38,316

	M$	Mills/kWh
First-year annual revenue requirements	26.97	9.81
Levelized annual revenue requirements	38.32	13.93

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-15. LIMESTONE PROCESS, CASE 31,
WITHOUT FORCED OXIDATION AND WITH ADDITIVE
SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, adipic acid, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,605
Feed preparation	4,306
Gas handling	10,517
SO_2 absorption	17,747
Reheat	3,223
Solids separation	2,472
Total process capital	40,870
Services, utilities, and miscellaneous	2,452
Total direct investment excluding pond	43,322
Pond construction	13,103
Total direct investment	56,425
Indirect Investment	
Engineering design and supervision	3,033
Architect and engineering contractor	866
Construction expense	6,932
Contractor fees	2,166
Contingency	5,632
Waste disposal indirect investment	3,610
Total fixed investment	78,664
Other Capital Investment	
Allowance for startup and modifications	4,956
Interest during construction	12,272
Royalties	0
Land	1,701
Working capital	3,111
Total capital investment	100,704
Dollars per kW of generation capacity	201.41

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-16. LIMESTONE PROCESS, CASE 31,

WITHOUT FORCED OXIDATION AND WITH ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, adipic acid, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	109.1 ktons	8.50/ton	927
Adipic acid	996.8 tons	1,500.00/ton	1,495
Total raw materials cost			2,422
Conversion costs			
Operating labor and supervision			
FGD	29,730 man-hr	15.00/man-hr	446
Utilities			
Steam	425,500 klb	2.50/klb	1,064
Process water	238,200 kgal	0.14/kgal	33
Electricity	38,821,000 kWh	0.037/kWh	1,436
Maintenance			
Labor and material			3,859
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			6,907
Total direct costs			9,329
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,625
Total first-year operating and maintenance costs			11,954
Levelized capital charges (14.7% of total capital investment)			14,803
Total first-year annual revenue requirements			26,757
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			22,545
Levelized capital charges (14.7% of total capital investment)			14,803
Levelized annual revenue requirements			37,348

	M$	Mills/kWh
First-year annual revenue requirements	26.76	9.73
Levelized annual revenue requirements	37.35	13.58

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-17. LIMESTONE PROCESS, CASE 32,
WITH FORCED OXIDATION AND ADDITIVE
SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, dibasic acid, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,636
Feed preparation	4,216
Gas handling	10,477
SO_2 absorption	18,448
Oxidation	1,623
Reheat	3,223
Solids separation	3,456
Total process capital	44,079
Services, utilities, and miscellaneous	2,645
Total direct investment excluding landfill	46,724
Landfill construction	3,362
Landfill equipment	890
Total direct investment	50,976
Indirect Investment	
Engineering design and supervision	3,271
Architect and engineering contractor	935
Construction expense	7,476
Contractor fees	2,336
Contingency	6,074
Waste disposal indirect investment	1,496
Total fixed investment	72,564
Other Capital Investment	
Allowance for startup and modifications	5,345
Interest during construction	11,320
Royalties	0
Land	548
Working capital	2,996
Total capital investment	92,773
Dollars per kW of generation capacity	185.55

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-18. LIMESTONE PROCESS, CASE 32,
WITH FORCED OXIDATION AND ADDITIVE
SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, dibasic acid, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	109.1 ktons	8.50/ton	927
Dibasic acid	189.8 tons	640.00/ton	121
Total raw materials cost			1,048
Conversion costs			
Operating labor and supervision			
FGD	42,920 man-hr	15.00/man-hr	644
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	425,200 klb	2.50/klb	1,063
Process water	178,300 kgal	0.14/kgal	25
Electricity	45,100,000 kWh	0.037/kWh	1,669
Diesel fuel	73,150 gal	1.60/gal	117
Maintenance			
Labor and material			3,866
Analysis	3,330 man-hr	21.00/man-hr	70
Total conversion costs			7,978
Total direct costs			9,026
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,062
Total first-year operating and maintenance costs			12,088
Levelized capital charges (14.7% of total capital investment)			13,638
Total first-year annual revenue requirements			25,726
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			22,797
Levelized capital charges (14.7% of total capital investment)			13,638
Levelized annual revenue requirements			36,435

	M$	Mills/kWh
First-year annual revenue requirements	25.73	9.35
Levelized annual revenue requirements	36.44	13.25

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-19. LIMESTONE PROCESS, CASE 33,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(1000-MW, 3.5% sulfur, 1979 NSPS, spray tower, adipic acid, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,917
Feed preparation	5,603
Gas handling	19,863
SO_2 absorption	33,988
Oxidation	4,417
Reheat	6,152
Solids separation	4,382
Total process capital	77,322
Services, utilities, and miscellaneous	4,639
Total direct investment excluding landfill	81,961
Landfill construction	5,221
Landfill equipment	1,466
Total direct investment	88,648
Indirect Investment	
Engineering design and supervision	4,918
Architect and engineering contractor	820
Construction expense	11,475
Contractor fees	3,278
Contingency	10,245
Waste disposal indirect investment	2,214
Total fixed investment	121,598
Other Capital Investment	
Allowance for startup and modifications	9,015
Interest during construction	18,969
Royalties	0
Land	847
Working capital	5,059
Total capital investment	155,488
Dollars per kW of generation capacity	155.49

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 2 spare absorber trains, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-20. LIMESTONE PROCESS, CASE 33,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(1000-MW, 3.5% sulfur, 1979 NSPS, spray tower, adipic acid, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	211.2 ktons	8.50/ton	1,796
Adipic acid	269 tons	1,500.00/ton	403
Total raw materials cost			2,199
Conversion costs			
Operating labor and supervision			
FGD	53,840 man-hr	15.00/man-hr	808
Waste disposal	37,440 man-hr	21.00/man-hr	786
Utilities			
Steam	823,500 klb	2.50/klb	2,059
Process water	345,300 kgal	0.14/kgal	48
Electricity	85,095,000 kWh	0.037/kWh	3,148
Diesel fuel	124,300 gal	1.60/gal	199
Maintenance			
Labor and material			5,938
Analysis	6,660 man-hr	21.00/man-hr	140
Total conversion costs			13,126
Total direct costs			15,325
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			4,603
Total first-year operating and maintenance costs			19,928
Levelized capital charges (14.7% of total capital investment)			22,857
Total first-year annual revenue requirements			42,785
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			37,584
Levelized capital charges (14.7% of total capital investment)			22,857
Levelized annual revenue requirements			60,441

	M$	Mills/kWh
First-year annual revenue requirements	42.79	7.78
Levelized annual revenue requirements	60.44	10.99

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-21. LIMESTONE PROCESS, CASE 34,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 0.7% sulfur, 1979 NSPS, spray tower, magnesium oxide, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,307
Feed preparation	2,380
Gas handling	11,020
SO_2 absorption	17,406
Oxidation	0
Reheat	0
Solids separation	2,275
Total process capital	34,388
Services, utilities, and miscellaneous	2,063
Total direct investment excluding landfill	36,451
Landfill construction	1,091
Landfill equipment	409
Total direct investment	37,951
Indirect Investment	
Engineering design and supervision	2,552
Architect and engineering contractor	729
Construction expense	5,832
Contractor fees	1,823
Contingency	4,739
Waste disposal indirect investment	509
Total fixed investment	54,135
Other Capital Investment	
Allowance for startup and modifications	4,170
Interest during construction	8,445
Royalties	0
Land	184
Working capital	2,165
Total capital investment	69,099
Dollars per kW of generation capacity	138.20

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 2 spare absorber trains, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-22. LIMESTONE PROCESS, CASE 34,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 0.7% sulfur, 1979 NSPS, spray tower,
magnesium oxide, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	20.3 ktons	8.50/ton	173
Magnesium oxide	18.3 tons	460.00/ton	8
Total raw materials cost			181
Conversion costs			
Operating labor and supervision			
FGD	39,620 man-hr	15.00/man-hr	594
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Process water	135,100 kgal	0.14/kgal	19
Electricity	37,160,000 kWh	0.037/kWh	1,375
Diesel fuel	36,260 gal	1.60/gal	58
Maintenance			
Labor and material			2,962
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			5,601
Total direct costs			5,782
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,489
Total first-year operating and maintenance costs			8,271
Levelized capital charges (14.7% of total capital investment)			10,158
Total first-year annual revenue requirements			18,429
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			15,599
Levelized capital charges (14.7% of total capital investment)			10,158
Levelized annual revenue requirements			25,757

	M$	Mills/kWh
First-year annual revenue requirements	18.43	6.70
Levelized annual revenue requirements	25.76	9.37

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-23. LIMESTONE PROCESS, CASE 35,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 2.0% sulfur, 1979 NSPS, spray tower, magnesium oxide, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,365
Feed preparation	3,094
Gas handling	10,742
SO_2 absorption	17,355
Oxidation	2,491
Reheat	3,606
Solids separation	3,012
Total process capital	41,665
Services, utilities, and miscellaneous	2,500
Total direct investment excluding landfill	44,165
Landfill construction	2,375
Landfill equipment	599
Total direct investment	47,139
Indirect Investment	
Engineering design and supervision	3,092
Architect and engineering contractor	883
Construction expense	7,066
Contractor fees	2,208
Contingency	5,741
Waste disposal indirect investment	1,051
Total fixed investment	67,180
Other Capital Investment	
Allowance for startup and modifications	5,052
Interest during construction	10,480
Royalties	0
Land	390
Working capital	2,810
Total capital investment	85,912
Dollars per kW of generation capacity	171.82

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-24. LIMESTONE PROCESS, CASE 35,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 2.0% sulfur, 1979 NSPS, spray tower, magnesium oxide, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	69.5 ktons	8.50/ton	591
Magnesium oxide	62.5 tons	460.00/ton	29
Total raw materials cost			620
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Waste disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	551,380 klb	2.50/klb	1,378
Process water	175,600 kgal	0.14/kgal	25
Electricity	38,160,000 kWh	0.037/kWh	1,412
Diesel fuel	48,340 gal	1.60/gal	77
Maintenance			
Labor and material			3,623
Analysis	4,960 man-hr	21.00/man-hr	104
Total conversion costs			7,801
Total direct costs			8,421
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,945
Total first-year operating and maintenance costs			11,366
Levelized capital charges (14.7% of total capital investment)			12,629
Total first-year annual revenue requirements			23,995
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			21,437
Levelized capital charges (14.7% of total capital investment)			12,629
Levelized annual revenue requirements			34,066

	M$	Mills/kWh
First-year annual revenue requirements	24.00	8.72
Levelized annual revenue requirements	34.07	12.39

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-25. LIMESTONE PROCESS, CASE 36,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower,
magnesium oxide, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,568
Feed preparation	4,689
Gas handling	11,213
SO_2 absorption	19,886
Oxidation	3,085
Reheat	3,627
Solids separation	3,695
Total process capital	48,763
Services, utilities, and miscellaneous	2,925
Total direct investment excluding landfill	51,688
Landfill construction	3,660
Landfill equipment	924
Total direct investment	56,272
Indirect Investment	
Engineering design and supervision	3,618
Architect and engineering contractor	1,034
Construction expense	8,270
Contractor fees	2,584
Contingency	6,719
Waste disposal indirect investment	1,620
Total fixed investment	80,117
Other Capital Investment	
Allowance for startup and modifications	5,913
Interest during construction	12,498
Royalties	0
Land	595
Working capital	3,363
Total capital investment	102,486
Dollars per kW of generation capacity	204.97

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-26. LIMESTONE PROCESS, CASE 36,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, magnesium oxide, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	132.6 ktons	8.50/ton	1,127
Magnesium oxide	119.4 tons	460.00/ton	55
Total raw materials cost			1,182
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	542,900 klb	2.50/klb	1,357
Process water	192,100 kgal	0.14/kgal	27
Electricity	55,537,000 kWh	0.037/kWh	2,055
Diesel fuel	77,040 gal	1.60/gal	123
Maintenance			
Labor and material			4,273
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			9,209
Total direct costs			10,391
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,388
Total first-year operating and maintenance costs			13,779
Levelized capital charges (14.7% of total capital investment)			15,066
Total first-year annual revenue requirements			28,845
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			25,988
Levelized capital charges (14.7% of total capital investment)			15,066
Levelized annual revenue requirements			41,054

	M$	Mills/kWh
First-year annual revenue requirements	28.85	10.49
Levelized annual revenue requirements	41.05	14.93

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-27. LIMESTONE PROCESS, CASE 37,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 5.0% sulfur, 1979 NSPS, spray tower, magnesium oxide, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,694
Feed preparation	5,146
Gas handling	11,147
SO_2 absorption	19,566
Oxidation	3,902
Reheat	3,592
Solids separation	4,227
Total process capital	50,274
Services, utilities, and miscellaneous	3,016
Total direct investment excluding landfill	53,290
Landfill construction	4,653
Landfill equipment	1,373
Total direct investment	59,316
Indirect Investment	
Engineering design and supervision	3,730
Architect and engineering contractor	1,066
Construction expense	8,526
Contractor fees	2,665
Contingency	6,928
Waste disposal indirect investment	2,098
Total fixed investment	84,329
Other Capital Investment	
Allowance for startup and modifications	6,096
Interest during construction	13,155
Royalties	0
Land	753
Working capital	3,570
Total capital investment	107,903
Dollars per kW of generation capacity	215.81

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-28. LIMESTONE PROCESS, CASE 37,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 5.0% sulfur, 1979 NSPS, spray tower,
magnesium oxide, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	189.7 ktons	8.50/ton	1,612
Magnesium oxide	170.7 tons	460.00/ton	79
Total raw materials cost			1,691
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Waste disposal	33,280 man-hr	21.00/man-hr	699
Utilities			
Steam	535,200 klb	2.50/klb	1,338
Process water	206,400 kgal	0.14/kgal	29
Electricity	59,106,000 kWh	0.037/kWh	2,187
Diesel fuel	119,800 gal	1.60/gal	192
Maintenance			
Labor and material			4,443
Analysis	5,000 man-hr	21.00/man-hr	105
Total conversion costs			9,651
Total direct costs			11,342
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,543
Total first-year operating and maintenance costs			14,885
Levelized capital charges (14.7% of total capital investment)			15,862
Total first-year annual revenue requirements			30,747
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			28,074
Levelized capital charges (14.7% of total capital investment)			15,862
Levelized annual revenue requirements			43,936

	M$	Mills/kWh
First-year annual revenue requirements	30.75	11.18
Levelized annual revenue requirements	43.94	15.98

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-29. LIMESTONE PROCESS, CASE 38, WITH FORCED OXIDATION AND ADDITIVE SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, TCA, magnesium oxide, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,574
Feed preparation	4,683
Gas handling	12,053
SO_2 absorption	17,006
Oxidation	3,097
Reheat	3,473
Solids separation	3,665
Total process capital	46,551
Services, utilities, and miscellaneous	2,793
Total direct investment excluding landfill	49,344
Landfill construction	3,660
Landfill equipment	924
Total direct investment	53,928
Indirect Investment	
Engineering design and supervision	3,454
Architect and engineering contractor	987
Construction expense	7,895
Contractor fees	2,467
Contingency	6,415
Waste disposal indirect investment	1,620
Total fixed investment	76,766
Other Capital Investment	
Allowance for startup and modifications	5,645
Interest during construction	11,975
Royalties	0
Land	595
Working capital	3,232
Total capital investment	98,213
Dollars per kW of generation capacity	196.43

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.

Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.

Reheat to 175°F as required.

Capital costs begin with an inlet plenum downstream of the ESP.

Stack plenum and stack excluded.

Disposal site one mile from power plant.

TABLE C-30. LIMESTONE PROCESS, CASE 38, WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, TCA,
magnesium oxide, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	132.6 ktons	8.50/ton	1,127
Magnesium oxide	119.4 tons	460.00/ton	55
Total raw materials cost			1,182
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	506,800 klb	2.50/klb	1,267
Process water	192,100 kgal	0.14/kgal	27
Electricity	53,433,000 kWh	0.037/kWh	1,977
Diesel fuel	77,040 gal	1.60/gal	123
Maintenance			
Labor and material			4,085
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			8,853
Total direct costs			10,035
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,276
Total first-year operating and maintenance costs			13,311
Levelized capital charges (14.7% of total capital investment)			14,437
Total first-year annual revenue requirements			27,748
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			25,105
Levelized capital charges (14.7% of total capital investment)			14,437
Levelized annual revenue requirements			39,542

	M$	Mills/kWh
First-year annual revenue requirements	27.75	10.09
Levelized annual revenue requirements	39.54	14.38

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-31. LIMESTONE PROCESS, CASE 39,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, magnesium oxide, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,568
Feed preparation	4,631
Gas handling	13,486
SO_2 absorption	23,335
Oxidation	3,086
Reheat	3,530
Solids separation	3,660
Total process capital	54,296
Services, utilities, and miscellaneous	3,258
Total direct investment excluding landfill	57,554
Landfill construction	3,598
Landfill equipment	916
Total direct investment	62,068
Indirect Investment	
Engineering design and supervision	4,029
Architect and engineering contractor	1,151
Construction expense	9,209
Contractor fees	2,878
Contingency	7,482
Waste disposal indirect investment	1,593
Total fixed investment	88,410
Other Capital Investment	
Allowance for startup and modifications	6,584
Interest during construction	13,792
Royalties	0
Land	582
Working capital	3,690
Total capital investment	113,058
Dollars per kW of generation capacity	226.12

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-32. LIMESTONE PROCESS, CASE 39,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, magnesium oxide, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	127.5 ktons	8.50/ton	1,084
Magnesium oxide	114.8 tons	460.00/ton	53
Total raw materials cost			1,137
Conversion costs			
Operating labor and supervision			
FGD	45,510 man-hr	15.00/man-hr	683
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	493,700 klb	2.50/klb	1,234
Process water	191,400 kgal	0.14/kgal	27
Electricity	71,840,000 kWh	0.037/kWh	2,658
Diesel fuel	76,750 gal	1.60/gal	123
Maintenance			
Labor and material			4,741
Analysis	4,970 man-hr	21.00/man-hr	105
Total conversion costs			10,182
Total direct costs			11,319
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,683
Total first-year operating and maintenance costs			15,002
Levelized capital charges (14.7% of total capital investment)			16,620
Total first-year annual revenue requirements			31,622
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			28,290
Levelized capital charges (14.7% of total capital investment)			16,620
Levelized annual revenue requirements			44,910

	M$	Mills/kWh
First-year annual revenue requirements	31.62	11.50
Levelized annual revenue requirements	44.91	16.33

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE C-33. LIMESTONE PROCESS, CASE 40,

WITHOUT FORCED OXIDATION AND WITH ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, magnesium oxide, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,565
Feed preparation	4,679
Gas handling	11,312
SO_2 absorption	20,248
Reheat	3,631
Solids separation	2,602
Total process capital	45,037
Services, utilities, and miscellaneous	2,702
Total direct investment excluding pond	47,739
Pond construction	14,644
Total direct investment	62,383
Indirect Investment	
Engineering design and supervision	3,342
Architect and engineering contractor	955
Construction expense	7,638
Contractor fees	2,387
Contingency	6,206
Waste disposal indirect investment	4,042
Total fixed investment	86,953
Other Capital Investment	
Allowance for startup and modifications	5,461
Interest during construction	13,564
Royalties	0
Land	1,899
Working capital	3,339
Total capital investment	111,216
Dollars per kW of generation capacity	222.43

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, spare absorber trains, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE C-34. LIMESTONE PROCESS, CASE 40, WITHOUT FORCED OXIDATION AND WITH ADDITIVE

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower,
magnesium oxide, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	132.6 ktons	8.50/ton	1,127
Magnesium oxide	119.4 tons	460.00/ton	55
Total raw materials cost			1,182
Conversion costs			
Operating labor and supervision			
FGD	30,680 man-hr	15.00/man-hr	460
Utilities			
Steam	542,600 klb	2.50/klb	1,357
Process water	262,400 kgal	0.14/kgal	37
Electricity	49,860,000 kWh	0.037/kWh	1,845
Maintenance			
Labor and material			4,258
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			8,061
Total direct costs			9,243
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,893
Total first-year operating and maintenance costs			12,136
Levelized capital charges (14.7% of total capital investment)			16,349
Total first-year annual revenue requirements			28,485
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			22,889
Levelized capital charges (14.7% of total capital investment)			16,349
Levelized annual revenue requirements			39,238

	M$	Mills/kWh
First-year annual revenue requirements	28.49	10.36
Levelized annual revenue requirements	39.24	14.27

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

APPENDIX D

LIME PROCESSES

TABLE D-1. LIME PROCESS, CASE 41,
WITHOUT FORCED OXIDATION OR ADDITIVES
SUMMARY OF CAPITAL INVESTMENT

(500-MW, 0.7% sulfur, 1979 NSPS,
spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,129
Feed preparation	961
Gas handling	10,340
SO_2 absorption	14,419
Reheat	0
Solids separation	1,599
Total process capital	28,448
Services, utilities, and miscellaneous	1,706
Total direct investment excluding pond	30,154
Pond construction	3,512
Total direct investment	33,666
Indirect Investment	
Engineering design and supervision	2,111
Architect and engineering contractor	603
Construction expense	4,825
Contractor fees	1,508
Contingency	3,920
Waste disposal indirect investment	969
Total fixed investment	47,602
Other Capital Investment	
Allowance for startup and modifications	3,449
Interest during construction	7,426
Royalties	0
Land	478
Working capital	1,778
Total capital investment	60,733
Dollars per kW of generation capacity	121.47

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-2. LIME PROCESS, CASE 41, WITHOUT FORCED OXIDATION OR ADDITIVES SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 0.7% sulfur, 1979 NSPS,
spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	9.2 ktons	75.00/ton	690
Total raw materials cost			690
Conversion costs			
Operating labor and supervision			
FGD	26,430 man-hr	15.00/man-hr	396
Utilities			
Process water	147,100 kgal	0.14/kgal	21
Electricity	24,058,000 kWh	0.037/kWh	890
Maintenance			
Labor and material			2,518
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			3,894
Total direct costs			4,584
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			1,790
Total first-year operating and maintenance costs			6,374
Levelized capital charges (14.7% of total capital investment)			8,928
Total first-year annual revenue requirements			15,302
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			12,020
Levelized capital charges (14.7% of total capital investment)			8,928
Levelized annual revenue requirements			20,948

	M$	Mills/kWh
First-year annual revenue requirements	15.30	5.56
Levelized annual revenue requirements	20.95	7.62

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-3. LIME PROCESS, CASE 42,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 2.0% sulfur, 1979 NSPS,
spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	1,901
Feed preparation	1,421
Gas handling	9,823
SO_2 absorption	16,258
Reheat	2,934
Solids separation	2,030
Total process capital	34,367
Services, utilities, and miscellaneous	2,062
Total direct investment excluding pond	36,429
Pond construction	7,735
Total direct investment	44,164
Indirect Investment	
Engineering design and supervision	2,550
Architect and engineering contractor	729
Construction expense	5,829
Contractor fees	1,821
Contingency	4,735
Waste disposal indirect investment	2,135
Total fixed investment	61,963
Other Capital Investment	
Allowance for startup and modifications	4,168
Interest during construction	9,666
Royalties	0
Land	1,020
Working capital	2,491
Total capital investment	79,308
Dollars per kW of generation capacity	158.62

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-4. LIME PROCESS, CASE 42,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 2.0% sulfur, 1979 NSPS, spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	31.3 ktons	75.00/ton	2,348
Total raw materials cost			2,348
Conversion costs			
Operating labor and supervision			
FGD	29,110 man-hr	15.00/man-hr	437
Utilities			
Steam	352,900 klb	2.50/klb	882
Process water	187,000 kgal	0.14/kgal	26
Electricity	27,301,000 kWh	0.037/kWh	1,010
Maintenance			
Labor and material			3,146
Analysis	3,300 man-hr	21.00/man-hr	69
Total conversion costs			5,570
Total direct costs			7,918
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,191
Total first-year operating and maintenance costs			10,109
Levelized capital charges (14.7% of total capital investment)			11,658
Total first-year annual revenue requirements			21,767
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			19,073
Levelized capital charges (14.7% of total capital investment)			11,658
Levelized annual revenue requirements			30,731

	M$	Mills/kWh
First-year annual revenue requirements	21.77	7.92
Levelized annual revenue requirements	30.73	11.18

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-5. LIME PROCESS, CASE 43,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS,
spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,395
Feed preparation	1,713
Gas handling	10,681
SO_2 absorption	17,879
Reheat	3,569
Solids separation	2,449
Total process capital	38,686
Services, utilities, and miscellaneous	2,321
Total direct investment excluding pond	41,007
Pond construction	12,785
Total direct investment	53,792
Indirect Investment	
Engineering design and supervision	2,871
Architect and engineering contractor	820
Construction expense	6,561
Contractor fees	2,050
Contingency	5,331
Waste disposal indirect investment	3,529
Total fixed investment	74,954
Other Capital Investment	
Allowance for startup and modifications	4,691
Interest during construction	11,693
Royalties	0
Land	1,668
Working capital	3,161
Total capital investment	96,167
Dollars per kW of generation capacity	192.33

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-6. LIME PROCESS, CASE 43, WITHOUT FORCED OXIDATION OR ADDITIVES SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS,
spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	60.0 ktons	75.00/ton	4,500
Total raw materials cost			4,500
Conversion costs			
Operating labor and supervision			
FGD	30,680 man-hr	15.00/man-hr	460
Utilities			
Steam	545,100 klb	2.50/klb	1,363
Process water	246,000 kgal	0.14/kgal	34
Electricity	32,929,000 kWh	0.037/kWh	1,218
Maintenance			
Labor and material			3,664
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			6,843
Total direct costs			11,343
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,537
Total first-year operating and maintenance costs			13,880
Levelized capital charges (14.7% of total capital investment)			14,137
Total first-year annual revenue requirements			28,017
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			26,183
Levelized capital charges (14.7% of total capital investment)			14,137
Levelized annual revenue requirements			40,320

	M$	Mills/kWh
First-year annual revenue requirements	28.02	10.19
Levelized annual revenue requirements	40.32	14.66

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-7. LIME PROCESS, CASE 44, WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 5.0% sulfur, 1979 NSPS,
spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	3,146
Feed preparation	1,931
Gas handling	10,661
SO_2 absorption	17,878
Reheat	3,535
Solids separation	2,722
Total process capital	39,873
Services, utilities, and miscellaneous	2,392
Total direct investment excluding pond	42,265
Pond construction	16,854
Total direct investment	59,119
Indirect Investment	
Engineering design and supervision	2,959
Architect and engineering contractor	845
Construction expense	6,762
Contractor fees	2,113
Contingency	5,495
Waste disposal indirect investment	4,652
Total fixed investment	81,945
Other Capital Investment	
Allowance for startup and modifications	4,836
Interest during construction	12,783
Royalties	0
Land	2,188
Working capital	3,526
Total capital investment	105,278
Dollars per kW of generation capacity	210.56

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-8. LIME PROCESS, CASE 44,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 5.0% sulfur, 1979 NSPS,
spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	85.8 ktons	75.00/ton	6,435
Total raw materials cost			6,435
Conversion costs			
Operating labor and supervision			
FGD	30,680 man-hr	15.00/man-hr	460
Utilities			
Steam	536,300 klb	2.50/klb	1,341
Process water	282,300 kgal	0.14/kgal	40
Electricity	33,172,000 kWh	0.037/kWh	1,227
Maintenance			
Labor and material			3,887
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			7,059
Total direct costs			13,494
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,670
Total first-year operating and maintenance costs			16,164
Levelized capital charges (14.7% of total capital investment)			15,476
Total first-year annual revenue requirements			31,640
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			30,490
Levelized capital charges (14.7% of total capital investment)			15,476
Levelized annual revenue requirements			45,966

	M$	Mills/kWh
First-year annual revenue requirements	31.64	11.51
Levelized annual revenue requirements	45.97	16.71

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-9. LIME PROCESS, CASE 45, WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS,
TCA, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,395
Feed preparation	1,713
Gas handling	11,434
SO_2 absorption	16,436
Reheat	3,454
Solids separation	2,433
Total process capital	37,865
Services, utilities, and miscellaneous	2,272
Total direct investment excluding pond	40,137
Pond construction	12,785
Total direct investment	52,922
Indirect Investment	
Engineering design and supervision	2,810
Architect and engineering contractor	803
Construction expense	6,422
Contractor fees	2,007
Contingency	5,216
Waste disposal indirect investment	3,529
Total fixed investment	73,709
Other Capital Investment	
Allowance for startup and modifications	4,591
Interest during construction	11,499
Royalties	0
Land	1,668
Working capital	3,130
Total capital investment	94,597
Dollars per kW of generation capacity	189.19

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-10. LIME PROCESS, CASE 45,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS,
TCA, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	60.0 ktons	75.00/ton	4,500
Total raw materials cost			4,500
Conversion costs			
Operating labor and supervision			
FGD	30,680 man-hr	15.00/man-hr	460
Utilities			
Steam	511,200 klb	2.50/klb	1,278
Process water	246,000 kgal	0.14/kgal	34
Electricity	37,109,000 kWh	0.037/kWh	1,373
Maintenance			
Labor and material			3,595
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			6,844
Total direct costs			11,344
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,495
Total first-year operating and maintenance costs			13,839
Levelized capital charges (14.7% of total capital investment)			13,906
Total first-year annual revenue requirements			27,745
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			26,105
Levelized capital charges (14.7% of total capital investment)			13,906
Levelized annual revenue requirements			40,011

	M$	Mills/kWh
First-year annual revenue requirements	27.75	10.09
Levelized annual revenue requirements	40.01	14.55

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-11. LIME PROCESS, CASE 46,

WITHOUT FORCED OXIDATION OR ADDITIVES

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS,
venturi - spray tower, pond disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,393
Feed preparation	1,713
Gas handling	12,489
SO_2 absorption	19,799
Reheat	3,470
Solids separation	2,434
Total process capital	42,298
Services, utilities, and miscellaneous	2,538
Total direct investment excluding pond	44,836
Pond construction	12,785
Total direct investment	57,621
Indirect Investment	
Engineering design and supervision	3,139
Architect and engineering contractor	897
Construction expense	7,174
Contractor fees	2,242
Contingency	5,827
Waste disposal indirect investment	3,529
Total fixed investment	80,429
Other Capital Investment	
Allowance for startup and modifications	5,128
Interest during construction	12,547
Royalties	0
Land	1,668
Working capital	3,384
Total capital investment	103,156
Dollars per kW of generation capacity	206.31

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-12. LIME PROCESS, CASE 46, WITHOUT FORCED OXIDATION OR ADDITIVES SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS,
venturi - spray tower, pond disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	60.0 ktons	75.00/ton	4,500
Total raw materials cost			4,500
Conversion costs			
Operating labor and supervision			
FGD	32,320 man-hr	15.00/man-hr	485
Utilities			
Steam	497,900 klb	2.50/klb	1,245
Process water	246,000 kgal	0.14/kgal	34
Electricity	45,166,000 kWh	0.037/kWh	1,671
Maintenance			
Labor and material			3,970
Analysis	4,940 man-hr	21.00/man-hr	104
Total conversion costs			7,509
Total direct costs			12,009
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,736
Total first-year operating and maintenance costs			14,745
Levelized capital charges (14.7% of total capital investment)			15,164
Total first-year annual revenue requirements			29,909
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			27,809
Levelized capital charges (14.7% of total capital investment)			15,164
Levelized annual revenue requirements			42,973

	M$	Mills/kWh
First-year annual revenue requirements	29.91	10.88
Levelized annual revenue requirements	42.97	15.63

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-13. LIME PROCESS, CASE 47, WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS,
spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,397
Feed preparation	1,710
Gas handling	10,643
SO_2 absorption	17,854
Oxidation	2,374
Reheat	3,568
Solids separation	3,412
Total process capital	41,958
Services, utilities, and miscellaneous	2,517
Total direct investment excluding landfill	44,475
Landfill construction	3,308
Landfill equipment	884
Total direct investment	48,667
Indirect Investment	
Engineering design and supervision	3,113
Architect and engineering contractor	890
Construction expense	7,116
Contractor fees	2,224
Contingency	5,782
Waste disposal indirect investment	1,474
Total fixed investment	69,266
Other Capital Investment	
Allowance for startup and modifications	5,088
Interest during construction	10,805
Royalties	0
Land	543
Working capital	3,197
Total capital investment	88,899
Dollars per kW of generation capacity	177.80

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-14. LIME PROCESS, CASE 47,
WITH FORCED OXIDATION
SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS,
spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	60.0 ktons	75.00/ton	4,500
Total raw materials cost			4,500
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	544,800 klb	2.50/klb	1,362
Process water	187,800 kgal	0.14/kgal	26
Electricity	38,937,000 kWh	0.037/kWh	1,441
Diesel fuel	72,940 gal	1.60/gal	117
Maintenance			
Labor and material			3,684
Analysis	4,970 man-hr	21.00/man-hr	105
Total conversion costs			7,917
Total direct costs			12,417
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,982
Total first-year operating and maintenance costs			15,399
Levelized capital charges (14.7% of total capital investment)			13,068
Total first-year annual revenue requirements			28,467
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			29,045
Levelized capital charges (14.7% of total capital investment)			13,068
Levelized annual revenue requirements			42,113

	M$	Mills/kWh
First-year annual revenue requirements	28.47	10.35
Levelized annual revenue requirements	42.11	15.31

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-15. LIME PROCESS, CASE 48,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, TCA, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,397
Feed preparation	1,710
Gas handling	11,437
SO_2 absorption	16,452
Oxidation	1,730
Reheat	3,452
Solids separation	3,396
Total process capital	40,574
Services, utilities, and miscellaneous	2,434
Total direct investment excluding landfill	43,008
Landfill construction	3,308
Landfill equipment	884
Total direct investment	47,200
Indirect Investment	
Engineering design and supervision	3,011
Architect and engineering contractor	860
Construction expense	6,881
Contractor fees	2,150
Contingency	5,592
Waste disposal indirect investment	1,474
Total fixed investment	67,168
Other Capital Investment	
Allowance for startup and modifications	4,921
Interest during construction	10,478
Royalties	0
Land	542
Working capital	3,135
Total capital investment	86,244
Dollars per kW of generation capacity	172.49

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-16. LIME PROCESS, CASE 48,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS,
TCA, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	57.7 ktons	75.00/ton	4,500
Total raw materials cost			4,500
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	510,860 klb	2.50/klb	1,277
Process water	187,800 kgal	0.14/kgal	26
Electricity	42,443,000 kWh	0.037/kWh	1,570
Diesel fuel	72,930 gal	1.60/gal	117
Maintenance			
Labor and material			3,566
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			7,843
Total direct costs			12,343
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,912
Total first-year operating and maintenance costs			15,255
Levelized capital charges (14.7% of total capital investment)			12,678
Total first-year annual revenue requirements			27,933
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			28,776
Levelized capital charges (14.7% of total capital investment)			12,678
Levelized annual revenue requirements			41,454

	M$	Mills/kWh
First-year annual revenue requirements	27.93	10.16
Levelized annual revenue requirements	41.45	15.07

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-17. LIME PROCESS, CASE 49,

WITH FORCED OXIDATION

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,397
Feed preparation	1,710
Gas handling	12,452
SO_2 absorption	19,436
Oxidation	1,069
Reheat	3,463
Solids separation	3,396
Total process capital	43,923
Services, utilities, and miscellaneous	2,635
Total direct investment excluding landfill	46,558
Landfill construction	3,308
Landfill equipment	884
Total direct investment	50,750
Indirect Investment	
Engineering design and supervision	3,259
Architect and engineering contractor	931
Construction expense	7,449
Contractor fees	2,328
Contingency	6,052
Waste disposal indirect investment	1,474
Total fixed investment	72,243
Other Capital Investment	
Allowance for startup and modifications	5,326
Interest during construction	11,270
Royalties	0
Land	542
Working capital	3,333
Total capital investment	92,714
Dollars per kW of generation capacity	185.43

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-18. LIME PROCESS, CASE 49,

WITH FORCED OXIDATION

SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS,
venturi - spray tower, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	60.0 ktons	75.00/ton	4,500
Total raw materials cost			4,500
Conversion costs			
Operating labor and supervision			
FGD	45,510 man-hr	15.00/man-hr	683
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	497,600 klb	2.50/klb	1,244
Process water	187,800 kgal	0.14/kgal	26
Electricity	49,764,000 kWh	0.037/kWh	1,841
Diesel fuel	72,930 gal	1.60/gal	117
Maintenance			
Labor and material			3,850
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			8,390
Total direct costs			12,890
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,097
Total first-year operating and maintenance costs			15,987
Levelized capital charges (14.7% of total capital investment)			13,629
Total first-year annual revenue requirements			29,616
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			30,155
Levelized capital charges (14.7% of total capital investment)			13,629
Levelized annual revenue requirements			43,784

	M$	Mills/kWh
First-year annual revenue requirements	29.62	10.77
Levelized annual revenue requirements	43.78	15.92

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-19. LIME PROCESS, CASE 50,

WITH FORCED OXIDATION AND ADDITIVE

SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower,
magnesium oxide, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,452
Feed preparation	1,751
Gas handling	10,624
SO_2 absorption	18,834
Oxidation	1,586
Reheat	3,564
Solids separation	3,496
Total process capital	42,307
Services, utilities, and miscellaneous	2,538
Total direct investment excluding landfill	44,845
Landfill construction	3,311
Landfill equipment	884
Total direct investment	49,040
Indirect Investment	
Engineering design and supervision	3,139
Architect and engineering contractor	897
Construction expense	7,175
Contractor fees	2,242
Contingency	5,831
Waste disposal indirect investment	1,475
Total fixed investment	69,799
Other Capital Investment	
Allowance for startup and modifications	5,130
Interest during construction	10,889
Royalties	0
Land	543
Working capital	3,211
Total capital investment	89,572
Dollars per kW of generation capacity	179.14

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-20. LIME PROCESS, CASE 50,
WITH FORCED OXIDATION AND ADDITIVE
SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, spray tower, magnesium oxide, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	60.0 ktons	75.00/ton	4,500
Magnesium oxide	54.0 tons	460.00/ton	25
Total raw materials cost			4,525
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	545,300 klb	2.50/klb	1,363
Process water	187,800 kgal	0.14/kgal	26
Electricity	37,622,000 kWh	0.037/kWh	1,392
Diesel fuel	72,940 gal	1.60/gal	117
Maintenance			
Labor and material			3,714
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			7,899
Total direct costs			12,424
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,000
Total first-year operating and maintenance costs			15,424
Levelized capital charges (14.7% of total capital investment)			13,167
Total first-year annual revenue requirements			28,591
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			29,093
Levelized capital charges (14.7% of total capital investment)			13,167
Levelized annual revenue requirements			42,260

	M$	Mills/kWh
First-year annual revenue requirements	28.59	10.40
Levelized annual revenue requirements	42.26	15.37

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-21. LIME PROCESS, CASE 51,
WITH FORCED OXIDATION AND ADDITIVE
SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, TCA,
magnesium oxide, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,451
Feed preparation	1,752
Gas handling	10,938
SO_2 absorption	17,638
Oxidation	2,968
Reheat	3,495
Solids separation	3,484
Total process capital	42,726
Services, utilities, and miscellaneous	2,564
Total direct investment excluding landfill	45,290
Landfill construction	3,311
Landfill equipment	884
Total direct investment	49,485
Indirect Investment	
Engineering design and supervision	3,170
Architect and engineering contractor	906
Construction expense	7,246
Contractor fees	2,264
Contingency	5,888
Waste disposal indirect investment	1,475
Total fixed investment	70,434
Other Capital Investment	
Allowance for startup and modifications	5,182
Interest during construction	10,988
Royalties	0
Land	543
Working capital	3,228
Total capital investment	90,375
Dollars per kW of generation capacity	180.75

Basis: North central new power unit with 30-year life at 5500 hr/yr full-load operation.
Mid-1982 costs, 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas.
Reheat to 175°F as required.
Capital costs begin with an inlet plenum downstream of the ESP.
Stack plenum and stack excluded.
Disposal site one mile from power plant.

TABLE D-22. LIME PROCESS, CASE 51,
WITH FORCED OXIDATION AND ADDITIVE
SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, TCA, magnesium oxide, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	60.0 ktons	75.00/ton	4,500
Magnesium oxide	54.0 tons	460.00/ton	25
Total raw materials cost			4,525
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	527,800 klb	2.50/klb	1,320
Process water	187,800 kgal	0.14/kgal	26
Electricity	38,230,000 kWh	0.037/kWh	1,416
Diesel fuel	72,940 gal	1.60/gal	117
Maintenance			
Labor and material			3,749
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			7,915
Total direct costs			12,440
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,021
Total first-year operating and maintenance costs			15,461
Levelized capital charges (14.7% of total capital investment)			13,285
Total first-year annual revenue requirements			28,746
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			29,160
Levelized capital charges (14.7% of total capital investment)			13,285
Levelized annual revenue requirements			42,445

	M$	Mills/kWh
First-year annual revenue requirements	28.75	10.45
Levelized annual revenue requirements	42.45	15.43

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TABLE D-23. LIME PROCESS, CASE 52,
WITH FORCED OXIDATION AND ADDITIVE
SUMMARY OF CAPITAL INVESTMENT

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower
magnesium oxide, landfill disposal)

	Capital investment, k$
Direct Investment	
Material handling	2,399
Feed preparation	1,730
Gas handling	12,594
SO_2 absorption	19,166
Oxidation	3,289
Reheat	3,460
Solids separation	3,450
Total process capital	46,088
Services, utilities, and miscellaneous	2,765
Total direct investment excluding landfill	48,853
Landfill construction	3,260
Landfill equipment	878
Total direct investment	52,991
Indirect Investment	
Engineering design and supervision	3,420
Architect and engineering contractor	977
Construction expense	7,816
Contractor fees	2,443
Contingency	6,351
Waste disposal indirect investment	1,453
Total fixed investment	75,451
Other Capital Investment	
Allowance for startup and modifications	5,589
Interest during construction	11,770
Royalties	0
Land	533
Working capital	3,414
Total capital investment	96,757
Dollars of total capital per kW of generation capacity	193.51

Basis: Upper Midwest plant location represents project beginning mid-1980, ending mid-1983. Average cost basis for scaling, mid-1982.
Minimum in-process storage, redundant scrubber train, pumps are spared.
Disposal pond located one mile from power plant.
FGD process investment begins at boiler air heater exit. Boiler plenum and stack excluded.
Only nominal construction overtime included.

TABLE D-24. LIME PROCESS, CASE 52, WITH FORCED OXIDATION AND ADDITIVE SUMMARY OF ANNUAL REVENUE REQUIREMENTS

(500-MW, 3.5% sulfur, 1979 NSPS, venturi - spray tower, magnesium oxide, landfill disposal)

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Lime	57.7 ktons	75.00/ton	4,328
Magnesium oxide	52.0 tons	460.00/ton	24
Total raw materials cost			4,352
Conversion costs			
Operating labor and supervision			
FGD	45,510 man-hr	15.00/man-hr	683
Solids disposal	24,960 man-hr	21.00/man-hr	524
Utilities			
Steam	498,800 klb	2.50/klb	1,247
Process water	187,300 kgal	0.14/kgal	26
Electricity	47,480,000 kWh	0.037/kWh	1,757
Diesel fuel	72,730 gal	1.60/gal	116
Maintenance			
Labor and material			4,032
Analysis	4,980 man-hr	21.00/man-hr	105
Total conversion costs			8,490
Total direct costs			12,842
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,206
Total first-year operating and maintenance costs			16,048
Levelized capital charges (14.7% of total capital investment)			14,223
Total first-year annual revenue requirements			30,271
Levelized first-year operating and maintenance costs (1.886 times first-year O&M)			30,267
Levelized capital charges (14.7% of total capital investment)			14,223
Levelized annual revenue requirements			44,490

	M$	Mills/kWh
First-year annual revenue requirements	30.27	11.01
Levelized annual revenue requirements	44.49	16.18

Basis: One-year, 5500-hour operation of system as described in capital investment table.
Mid-1984 cost basis.

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/7-83-029	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Economic Evaluation of Limestone and Lime Flue Gas Desulfurization Processes		5. REPORT DATE May 1983 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) T.A. Burnett, C.D. Stephenson, F.A. Sudhoff, and J.D. Veitch		8. PERFORMING ORGANIZATION REPORT NO. TVA/OP/EDT-82/6
9. PERFORMING ORGANIZATION NAME AND ADDRESS TVA, Office of Power Division of Energy Demonstration and Technology Muscle Shoals, Alabama 35660		10. PROGRAM ELEMENT NO. 11. CONTRACT/GRANT NO. EPA IAG-79-D-X0511
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Final; 8/79 - 1/83 14. SPONSORING AGENCY CODE EPA/600/13

15. SUPPLEMENTARY NOTES IERL-RTP project officer is J. David Mobley, MD-61, 919/541-2578.

16. ABSTRACT The preliminary-grade economics (accuracy: -15%, +30%) of various alternative limestone scrubbing options (absorber type, with and without forced oxidation, and with and without adipic acid enhancement) are examined using the current design and economic premises established for the continuing series of economic evaluations performed by TVA for EPA. The economics are projected using the Shawnee lime/limestone computer model, based on long-term operating data from EPA's alkali scrubbing test facility at TVA's Shawnee steam plant. The capital investment for the base-case limestone scrubbing process (500 MW, 3.5% sulfur coal, 1979 NSPS, spray tower, forced oxidation, landfill) is $206/kW. The first-year and levelized annual revenue requirements are 10.59 and 15.09 mills/kWh, respectively. Costs for the equivalent limestone scrubbing process using a turbulent contact absorber are lower, while those for the venturi/spray tower absorber are higher. Forced-oxidation/landfill disposal has a lower capital investment than unoxidized/pond disposal for all options studied; however, the first-year and levelized annual revenue requirements are slightly higher for forced-oxidation/landfill disposal for most coal applications. Use of adipic acid remains economically attractive even assuming both a high unit cost and a high degradation factor for adipic acid.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Scrubbers	Pollution Control	13B	13I
Flue Gases	Coal	Stationary Sources	21B	21D
Desulfurization	Adipic Acid		07A,07D	07C
Calcium Carbonates			07B	
Calcium Oxides				
Economic Analysis			05C	

18. DISTRIBUTION STATEMENT Release to Public	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 325
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA Form 2220-1 (9-73)